University of Alberta

Performance Evaluation of a Full-scale Ultrafiltration Membrane Used for Tertiary Treatment of Municipal Wastewater

by

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Abstract

A full-size ZeeWeed® 500D membrane system was operated as a tertiary treatment at GoldBar Wastewater Treatment Plant. During the ten months of operation, the impact of different feed waters and membrane operating conditions on permeate water quality and membrane performance has been investigated.

The permeate contained less than 0.6 mg/L total suspended solids, 0.2 NTU, 2 mg/L 5-day biochemical oxygen demand (BOD_5), low coliform and coliphage content regardless of the feed water and membrane operation. Silt density index (SDI) was used to prove that the permeate water quality has certain relationship with the membrane operating condition.

For membrane operating conditions, the impact of aeration and water recovery was negligible; backpulse duration was selected as 30 seconds; the optimized operating parameter range was between 16.5 and 33.4 $Lm^{-2}h^{-1}$ for flux and from 10 to 25 minutes for operating time.

NaOCl and citric acid maintenance chemical cleaning was investigated. NaOCl recovery cleaning reached highest cleaning effectiveness.

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Table of Contents

1. INTRODUCTION 1

2. BACKGROUND AND THEORY 3

2.1 Membrane Technology 3
2.1.1 Membrane Definitions 3
2.1.2 Membrane Modules and Critical Flux 3
2.1.3 Classification of Membrane 5
2.1.4 Limitation of Membrane Technology 7
2.1.5 Membrane Cleaning 9

2.2 Application of Membrane in Wastewater Reuse 10
2.2.1 Wastewater Reclamation and Reuse 10
2.2.1.1 Concerns of Water Reuse in Agriculture 11
2.2.1.2 Concerns of Water Reuse in Industry 15
2.2.2 Application of UF and MF Membrane in Wastewater Reuse 17
2.2.3 Application of NF and RO Membrane in Wastewater Reuse 19

2.3 Municipal Wastewater Plant Process and Water Quality 20
2.3.1 Conventional Secondary Treatment 20
2.3.2 The Biological Nutrient Removal Process 21
2.3.2.1 Biological Phosphorus Reduction 21
2.3.2.2 Biological Nitrogen Reduction (BNR) 24
2.3.2.2.1 Nitrification 25
2.3.2.2.2 Denitrification 26
2.3.3 Membrane Bioreactor 28

2.4 Gold Bar Wastewater Treatment Plant 30

3. METHODS AND MATERIALS 32

3.1 Experimental Equipment and Control 32
3.1.1 Membrane System 32
3.1.2 ZeeWeed Ultrafiltration Membrane 35
3.1.3 Production Process 37
3.1.4 Chemical Cleaning 39
3.1.4.1 Maintenance Cleaning 40
3.1.4.2 Recovery Cleaning 41

3.1.5 Parameters for Membrane Performance 42

3.2 Test Runs 47
3.2.1 Bubble Test 47
3.2.2 Tracer Test 49

3.3 Experiment Design 51

3.4 Sampling Design 52

3.5 Analytical Methods 54
3.5.1 TSS Determination 54
3.5.2 BOD_5 Determination 55
3.5.3 COD Determination 55
3.5.4 Ammonia Determination 55
3.5.5 Total Oxidized Nitrogen Determination 56
3.5.6 Total Kjeldahl Nitrogen Determination 56
3.5.7 Total Phosphorus Determination 56
3.5.8 Turbidity 56
3.5.9 pH 56
3.5.10 Coliforms 57
3.5.11 Coliphage 57
3.5.12 Metals Determination 58
3.5.13 Particle Count 58
3.5.14 Silt Density Index 58
3.5.15 Scanning Electron Microscopy 59

4. RESULTS AND DISCUSSION 60

4.1 Permeate Water Quality 60
4.1.1 Solids Removal 64
4.1.2 COD Removal 65
4.1.3 Nitrogen Removal 66
4.1.4 Phosphorus Removal 68
4.1.5 Metal Removal 68
4.1.6 Microorganism Reduction 69
4.1.7 Coliphage Test 75
4.1.8 Silt Density Index 77

4.2 Operating Factors 83
4.2.1 Flux 83
4.2.2 Operating time 86

4.2.3 Duration of Backpulse 88
4.2.4 Aeration 91
4.2.5 Water Recovery 93
4.2.6 Impact of Different Feed Water 94

4.3 Membrane Operating Parameters Range 96

4.4 Maintenance and Recovery Cleaning Effectiveness 97

4.5 Membrane Fouling 101

4.6 Others 107
4.6.1 Aeration and Backpulse Cleaning 107
4.6.2 Foaming 109
4.6.3 Energy Input 109
4.6.4 Membrane Module Appearance 110
4.6.5 Utility Water Used as Feed 113

5 CONCLUSIONS 114

6 RECOMMENDATIONS 117

REFERENCE: 118

APPENDIX A: CONTROL PROCEDURE OF STANDBY, BACKPULSE, PRODUCTION AND SHUTDOWN 123

APPENDIX B: MAINTENANCE CLEANING PROCEDURE 125

APPENDIX C: RECOVERY CLEANING PROCEDURE 126

APPENDIX D: SUMMARY OF OPERATION RECORD 128

APPENDIX E: WATER QUALITY RAW DATA 132

APPENDIX F: CALCULATION PROCEDURE FOR FIGURE 4.30. 143

List of Tables

Table 2.1 Characteristics of filtration processes driven by pressure (Cheryan, 1998). 7

Table 2.2 Summary of major parameters used to characterize reclaimed wastewater quality 12

Table 2.3 Comparison of criteria (maximum limits) for the irrigation of crops consumed by humans with reused wastewater by WHO, U.S. EPA, the State of California and some Mediterranean countries (national guidelines) (Angelakis et al., 1999). 13

Table 3.1 Characteristics of the UF unit. 35

Table 3.2 Chemicals required for cleaning the ZeeWeed system. 40

Table 3.3 Operating parameters for the tracer test. 49

Table 3.4 Measured and theoretical hydraulic retention time for pilot membrane plant. 51

Table 3.5 Summary of the operating parameters set points for FE after UV used as feed. 53

Table 3.6 Summary of the operating parameters set points for FE before UV used as feed. 54

Table 4.1 The summary of FE1 (influent after UV) and permeate water qualities. 62

Table 4.2 The summary of FE2 (influent before UV) and permeate water qualities. 63

Table 4.3 The reduction of metals. 69

Table 4.4 Summary of coliform in the membrane tank. 69

Table 4.5 Comparison of aerosol coliform inside and outside of the membrane tank. 71

Table 4.6 Coliphage in permeate and influent water samples 75

Table 4.7 The comparison of different cleaning efficiency. 100

List of Figures

Figure 2.1 Dead-end and cross-flow module. 5

Figure 2.2 Filtration processes of pressure membrane. 6

Figure 2.3 Biological phosphorus removal process. 23

Figure 2.4 Schematic flow diagrams for the membrane bioreactor activated sludge process: (a) with internal membrane biosolids separation unit and (b) with external biosolids separation unit. 28

Figure 2.5 GBWWTP full-scale BNR process. 31

Figure 3.1 ZeeWeed 500D membrane pilot plant in GBWWTP. 32

Figure 3.2 Experimental set-ups. 33

Figure 3.3 Human machine interface (HMI). 34

Figure 3.4 The schematic of ZeeWeed 500 module. 36

Figure 3.5 Membrane cassette used for this project. 37

Figure 3.6 Production process. 38

Figure 3.7 Backpulse process. 39

Figure 3.8 TMP calculated point. 44

Figure 3.9 TMP versus time in operating cycles. 45

Figure 3.10 Permeate flux versus time in operating cycles. 45

Figure 3.11 The sodium ion concentration versus time on January 9. 50

Figure 3.12 The sodium ion concentration versus time on January 13. 51

Figure 4.1 Variation of TSS when FE1 used as the feed water. 64

Figure 4.2 Variation of TSS when FE2 used as the feed water. 65

Figure 4.3 Variation of COD when FE1 used as the feed water. 66

Figure 4.4 Variation of COD when FE2 used as the feed water. 66

Figure 4.5 Permeate and influent TC for FE1 as the feed water. 72

Figure 4.6 Permeate and influent FC for FE1 as the feed water. 72

Figure 4.7 Permeate and influent TC for FE2 as the feed water. 73

Figure 4.8 Permeate and influent FC for FE2 as the feed water. 73

Figure 4.9 Backpulse tank before chemical cleaning. 74

Figure 4.10 Backpulse tank after chemical cleaning. 74

Figure 4.11 Coliphage concentration against time in one operating cycle. 76

Figure 4.12 The effect of flux and reversible fouling on SDI. 79

Figure 4.13 TMP when the SDI samples collected. 79

Figure 4.14 SEM pictures of Cake layer on the membrane surface and solids sucked in the membrane pores. 80

Figure 4.15 SDI data versus long-term operation. 81

Figure 4.16 TMP versus long-term operation. 81

Figure 4.17 HTMP versus time under 15 minutes operating time. 85

Figure 4.18 HTMP versus time under 10 minutes operating time. 86

Figure 4.19 HTMP versus time under 25 minutes operating time. 86

Figure 4.20 HTMP versus time under flux 23.7 $L/m^2/h$ and BP 30 sec. 88

Figure 4.21 The impact of backpulse duration on specific flux. 90

Figure 4.22 Actual flux versus time in one operating cycle with the same flux set point but different backpulse duration. 90

Figure 4.23 TMP versus one operating cycle with the same flux set point but different backpulse duration. 91

Figure 4.24 Impact of different aeration amounts with the same other operating set points (operating time15 min, flux 23.7 $Lm^{-2}h^{-1}$, backpulse 30 seconds). 92

Figure 4.25 Impact of different aeration amounts with the same other operating set points (operating time 25 min, flux 14.3 $Lm^{-2}h^{-1}$, backpulse 30 seconds). 92

Figure 4.26 Impact of different aeration amounts with the same other operating set points (operation time 10 min, flux 28.4 $Lm^{-2}h^{-1}$, backpulse 30 seconds). 93

Figure 4.27 HTMP versus time with different water recovery. 94

Figure 4.28 HTMP versus time with different feeds (operating time 10 minutes, backpulse 30 seconds, flux 23.7 $Lm^{-2}h^{-1}$, aeration 237 m^3h^{-1}, water recovery 70%) 95

Figure 4.29 HTMP versus time with different feeds (operating time 15 minutes, backpulse 30 seconds, flux 23.7 $lm^{-2}h^{-1}$, aeration 237 m^3h^{-1}, water recovery 70%) 95

Figure 4.30 Flux versus TMP with different operating time. 97

Figure 4.31 Specific flux after different chemical cleaning. 101

Figure 4.32 Specific flux versus time under 10 minutes operating time, 30 seconds BP, 28.4 $Lm^{-2}h^{-1}$ flux, 237 m^3h^{-1} aeration. 104

Figure 4.33 Specific flux versus time under 15 minutes operating time, 30 seconds BP, 23.4 $Lm^{-2}h^{-1}$ flux, 237 m^3h^{-1} aeration. 104

Figure 4.34 Specific flux versus time under 25 minutes operating time, 30 seconds BP, 23.7 $Lm^{-2}h^{-1}$ flux, 237 m^3h^{-1} aeration. 105

Figure 4.35 Specific flux at certain time. 105

Figure 4.36 Specific flux versus time under 25 minutes operating time, 30 seconds BP, 19.0 $Lm^{-2}h^{-1}$ flux, 237 m^3h^{-1} aeration. 106

Figure 4.37 Specific flux at certain time. 106

Figure 4.38 Specific flux versus time after a change from high flux to low flux. 107

Figure 4.39 TMP before and after 5 minutes shutdown. 108

Figure 4.40 TMP before and after 10 minutes shutdown. 108

Figure 4.41 Foaming during NaOCl maintenance cleaning. 109

Figure 4.42 Membrane appearance after one-month operation. 111

Figure 4.43 Membrane appearance after 10 months operation. 112

Figure 4.44 Snails taken from the membrane surface. 112

Figure 4.45 TMP against time when utility water used as the feed water. 113

List of Symbols

A	Pressure due to elevation of pressure transmitter
B	Distance from level transmitter to top of membrane
C	Elevation to pressure conversion factor
μ_T	Viscosity
T	Temperature
Γ	Hydraulic retention time
Q	Volume
V	Velocity
C_0	Original concentration
$T_{initial}$	Duration of time for initial 500 mL sample
T_{final}	Duration of time for final 500 mL sample

List of Abbreviations

BNR	Biological nutrient removal
BOD	Biochemical oxygen demand
BOD_5	5-day biochemical oxygen demand
BP	Backpulse
CFU	Colony forming units
COD	Chemical oxygen demand
EBPR	Enhance biological phosphorus removal
FC	Fecal coliform
FE1	Final effluent from GBWWTP that has been sterilized by UV
FE2	Final effluent from the tank #6 of GBWWTP without the UV sterilization
GBWWTP	Gold Bar wastewater treatment plant
HMI	Human machine interface
HTMP	Highest transmembrane pressure
MBR	Membrane bioreactors
MF	Microfiltration
MLE	Modified Luzack-Ettinger
MLSS	Mixed liquor suspended solid
MWCO	Molecular weight cut-off

NF	Nanofiltration
NH_3H	Ammonia
NTU	Nephelometric turbidity units
PHB	Polyhydroxybutyrate
PLC	Programmable logic controller
RO	Reverse osmosis
SAM	Sequencing anoxic/anaerobic membrane
SDI	Silt density index
SEM	Scanning electron microscopy
TA	Total alkalinity
TC	Total coliform
TDS	Total dissolved solids
TH	Total hardness
TKN	Total Kjeldahl nitrogen
TMP	Transmembrane pressure
TOC	Total organic carbon
TON	Total oxidized nitrogen
TP	Total phosphorus
TSS	Total suspended solid
UF	Ultrafiltration
VFA	Volatile fatty acids

1. Introduction

As we move into the new millennium, increasing population and water use lead to increased demands on water resources worldwide. At the same time, the long-term trend of climate change towards warmer summer temperatures and drier winters will challenge conventional surface and groundwater resources. The investigation and use of alternative water sources where appropriate is expected.

Reclaimed municipal wastewater reuse has become a common practice in many places, while the regulatory requirements for reuse applications are becoming more and more stringent. In such a regulatory framework, ultrafiltration membrane technology can provide a basic technique for tertiary treatment as assistance meeting regulatory requirements. Ultrafiltration membrane technology offers a physical separation for solids, bacteria and many viruses and provides a high flux with a lower energy requirement than nanofiltration or reverse osmosis.

The Gold Bar Wastewater Treatment Plant (GBWWTP), Edmonton, Alberta, had converted to full biological nutrient removal treatment by the end of 2002. In the past two years, GBWWTP produced excellent quality effluent, which the surrounding industry has shown a great interest in as a water source.

A full-scale ZeeWeed® 500D ultrafiltration membrane was designed and constructed at the GBWWTP as a tertiary treatment to further polish the secondary effluents. Both final effluent before UV and final effluent after UV were fed to the membrane system. The permeate water quality and membrane performance were the main study factors of this project. The pilot plant success as a tertiary treatment option for the final effluent was used to direct design and scaling-up of a larger facility.

2. Background and Theory

2.1 Membrane Technology

2.1.1 Membrane Definitions

Membrane technology is devoted to the separation of the minute particles ranging from bacteria to molecules from a liquid, in this case wastewater (Cardew and Le, 1998). The primary role of a membrane is to act as a selective barrier: permitting passage of certain components to one stream called permeate and retain other materials in another stream called concentrate (Cheryan, 1998). Selectivity comes through the interaction between the membrane and the surrounding phase. Two factors contribute to the selectivity, the partitioning of molecules and/or particles between the membrane and the surrounding phase, and the relative diffusion rates of these materials in the membrane.

2.1.2 Membrane Modules and Critical Flux

Membrane separation approaches include dead-end module design and cross-flow module design (Figure 2.1). For the dead-end module design, all the influent will pass through the membrane and all the retained particles will accumulate on the membrane surface. For cross-flow system the fluid moves tangentially across the membrane surface and part of fluid called concentrate will remain with the most retained particles. The consequence of cross-flow is that in continuous operation the flux through the membrane tends to be constant while in dead-end the flux decreases with time (Cardew and Le, 1998).

Cross-flow systems prevent the deposit of a "cake layer" at the membrane surface. It has been found that no cake layer was deposited when the membrane is operated at the subcritical flux condition. The maximum flux in this condition is called the critical flux. Flux rates above the critical flux value result in a decline in flux; flux rates below the critical flux value do not develop a deposit on cake layer. The critical flux value depends on several variables such as the ratio between the particle and the pore size, velocity, the transmembrane pressure and cross-flow channel dimensions (Riln, 2004). Defrance and Jaffrin (1999) measured critical flux for different velocities and found that the critical flux value increases approximately linearly with velocity. A concentration polarization model was developed to predict the critical flux condition and was verified with experimental results (Kim and Park, 2002).

While the critical flux concept has proved an invaluable tool in conventional membrane process design, operation below the critical flux is limited for real world applications. This is because the limitation to increasing crossflow velocity and its sensitivity to the change of feed composition (Kim and Park, 2002). Also, its validity is questionable in MBR processes where fouling rates only approach zero at very low flux and, ultimately, impractical values (Chang et al., 2002).

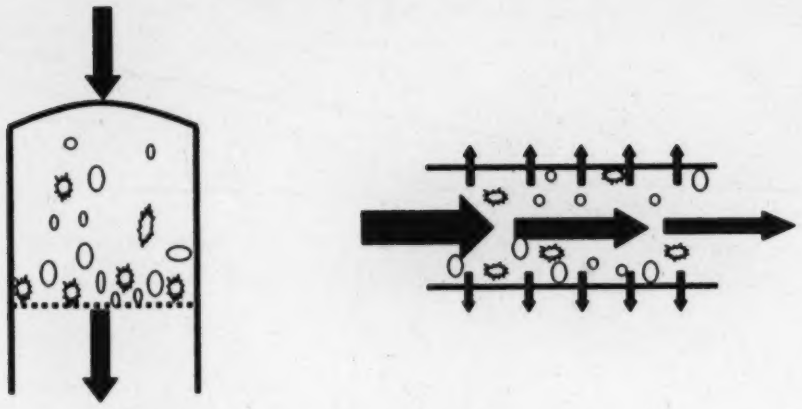

Figure 2.1. Dead-end and cross-flow module.

2.1.3 Classification of Membrane

Based on the driven forces, all the membrane processes can be defined into four categories: Electrically Driven Process, Thermally Driven Process, Concentration Driven Process and Pressure Driven Process. Electrodialysis, a main kind of electrically driven process, removes ionic material from a solution by using ion-selective membranes (Rautenbach and Albrecht, 1989). The major application of electrodialysis is to produce potable water from seawater and brackish waters. While pervaporation and membrane distillation are the samples of thermally driven process and they are unusual among membrane processes. Concentration driven process (dialysis), which is defined as the process that the flux of dissolved lower molecular mass components goes through the membrane as a result of a difference in trans-membrane concentration (Rautenbach & Albrecht, 1989), can be used in the medical community and can also be used to remove alcohol from beer.

The pressure driven process is divided in four types (See Table 2.1 and Figure 2.2). Reverse osmosis retains all components other than the solvent (e.g. water) itself, while ultrafiltration

retains only macromolecules or particles larger than about 10 to 200 A (about 0.001 to 0.02um). Microfiltration is designed to retain particles in the "micro" range, that is, suspended particles in the range of 0.10 μm to about 5 μm (particles larger than 5 to 10 μm are better separated using conventional cake filtration methods) (Cheryan, 1998). Thus, in its broadest sense, reverse osmosis is essentially considered to be a pure water production technique, while ultrafiltration can be looked as a method for simultaneously purifying, concentrating, and fractionating macromolecules or fine colloidal suspensions. Microfiltration is used mainly as a clarification technique, separating suspended particles from dissolved substance, provided the particles meet the size requirements for microfiltration membrane (Cheryan, 1998). Microfiltration membranes are characterized in terms of pore size, while ultrafiltration membranes are normally described in terms of a molecular weight cut-off (MWCO). However, there is no key principle separating one technology from another.

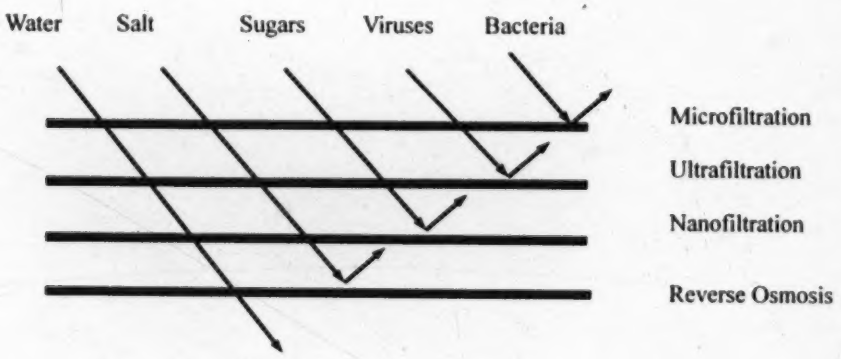

Figure 2.2. Filtration processes of pressure membrane.

Table 2.1. Characteristics of filtration processes driven by pressure (Cheryan, 1998).

Process Technology	Separation Principle	Size Range	MWCO	Rejection Characterization
Microfiltration	Size	0.1 μm to 1 μm	-	Absolute, nominal, or beta
Ultrofiltration	Size, Charge	1 nm to 100 nm	>1000	MWCO
Nanofiltration	Size, Charge, Affinity	~1 nm	200 to 1000	Rejection, MWCO
Reverse Osmosis	Size, Charge, Affinity	<1 nm	<200	Rejection

2.1.4 Limitation of Membrane Technology

In the late twenty century, membrane technology underwent major developments and began to be used in a number of fields: automotive, cosmetic, metal fabrication, food and beverage processing, landfill leachate, and other industries. But the main application of membranes is for separation. Compared with other separation techniques, the main advantages offered by membranes are (1) effluent properties for nonpotable reuse directly, (2) handling wide fluctuations in influent quality, and (3) reduced footprint (Delgado et al., 2004). Despite their inherent advantages, cost and fouling were the major barriers to the further development of membrane technology.

Costs have been reduced by developments of longer operating life, cheaper replacement and reduction of energy consumption (Marrot et al., 2004). Now membranes are enabling the affordable and cost effective reuse of wastewater as an alternative water resource (Reith and

Birkenhead, 1998). However, fouling resulting from the formation of a layer or cake on the membrane surface and/or preferential adsorption remains the main limitation for developing and expanding the use of membranes (Delgado et al., 2004). Fouling is indicated by a reduction of water flow through a given area of membrane at a given temperature, salt concentration and pressure. The materials that cause fouling may be biological in origin (bacteria), nonbiological (colloids, silts and clays) or chemical precipitation (scaling).

Many studies have focused on the optimization of membrane operating parameters to reduce membrane fouling (Ahn and Song, 2000; Fradin and Field, 1999). Permeate flux, suction time, aeration rate, concentration ratio and intermittent permeation were found to be the main factors for optimization of operating parameters in the membrane process (Chua et al., 2002; Liu and Wu, 1998, Gui et al., 2003). For membrane bioreactor, F/M ratio and solid retention time also are considered as major factors to impact membrane fouling (Cho *et al.*, 2003).

Chang *et al.* (2002) found that flux was the most significant factor in determining fouling rate. At high flux, colloidal aggregation and heterogeneous deposits were observed. Rapid reversible fouling then took place, predominantly through formation and compaction of the cake layer. Internal and/or irreversible fouling also took place more rapidly at higher fluxes. The critical flux concept, however, was challenged by Ahn and Song (2000) finding that membrane fouling occurred at sub-critical flux in the long-term operation. Therefore, chemical cleaning was necessary.

Another most common problem is scaling. Scaling is caused by a concentration of an inorganic salt in excess of its saturation point, which creates salt precipitation and deposition on or in the membrane and subsequently in the feed channel spaces (AWWA, 1991).

2.1.5 Membrane Cleaning

Membrane cleaning comprises intermittent physical cleaning (usually backwashing) and periodic chemical cleaning. Physical cleaning can produce a stable flux without secondary chemical contamination but is required more frequent and generally requires more energy.

Chemical cleaning is expected to completely recover membrane flux. Xing *et al.* (2003) found that the optimized chemical cleaning could restore the membrane's standard permeability to higher than 94% of a new membrane. It has also been found that chemical cleaning neither damaged the membrane nor affected filtration performance (Ahn and Song, 2000). But chemical cleaning produces toxic or contaminated wastewater because most cleaning agents are caustic and/or contain detergents and oxidizing agents such as hypochlorite. When microbiocides, particularly chlorine are used, they may be advantageous to operation but can also exacerbate biofouling problems. Microorganisms subjected to low levels of biocides often exude large amounts of extra cellular polysaccharides (EPS) as a protection, and it is this EPS material that forms the biofilm (Baker and Budley, 1998). Successful membrane cleaning procedures generally

employ some combination of physical cleaning and chemical cleaning (Chang *et al.*, 2002).

2.2 Application of Membrane in Wastewater Reuse

2.2.1 Wastewater Reclamation and Reuse

Continued population growth, contamination of both surface water and groundwater, uneven distributions of water resources, and periodic droughts have forced water agencies and the public to accept the wastewater reused as a water resource. However, water quality characterization is necessary to evaluate the biological and chemical safety to reuse reclaimed water for various applications.

The water quality parameters that are used to evaluate reclaimed wastewater are based on public and environmental health protection related to each type of water use. A summary of relevant water quality monitoring parameters is given in Table 2.2.

The nonpotable reuse mainly includes agricultural and landscape irrigation, and industrial water reuse. Most obvious benefit of this application is the reduction of demands on available surface and groundwater, and the reduction of water discharges, which eases the environmental impact (Strauss, 1991).

2.2.1.1 Concerns of Water Reuse in Agriculture

The use of reclaimed water for irrigation of landscaped areas and golf courses in the urban environment has recently been the largest wastewater reuse application. Due to the evaporation, the salt deposition from the applied wastewater will tend to accumulate in the soil and affect the physical and mechanical properties of the soil, such as degree of dispersion of the soil particles, stability of aggregates, soil structure, and permeability. At the same time, some special ions have specific toxicity to the plant growth, which is referred to as "specific ion toxicity". Thus, when the reclaimed water is being planned for irrigation, the physical and chemical characteristics are the most important concern (Angelakis et al., 1999). Secondly, nutrients will be another concern. Although nutrients are useful to agriculture and landscape management including N, P, and occasionally K, Zn, B, and S, excessive nitrogen in the crop growth will cause delayed or uneven maturity, or reduced crop quality. Also, clogging problems caused by biological growth and high concentrations of algae and suspended solids, have been reported (Metcalf & Eddy, 2003).

The guidelines for wastewater reuse through irrigation cover four areas: chemical standards, microbiological standards, wastewater treatment processes and irrigation techniques. A brief of these criteria is given in Table 2.3.

Table 2.2. Summary of major parameters used to characterize reclaimed wastewater quality.

Parameters	Significant in Wastewater Reclamation	Approximate Range in Treated Wastewater	Treatment Goal in Reclaimed Wastewater
Organic indicators			
BOD_5	Organic substrate for microbial or algal growth	10 to 30 mg/L	<1 to 10 mg/L
TOC	Measure of organic carbon	1 to 20 mg/L	<1 to 10 mg/L
Measurement of particulate matter			
TSS	Measure of particles in wastewater can be related to microbial contamination, turbidity can interfere with disinfections effectiveness.	<1 to 30 mg/L	<1 to 10 mg/L
Turbidity	Measure of particles in wastewater, can be related to TSS	1 to 30 NTU	0.1 to 10 NTU
Pathogenic organisms	Measure of microbial health risks due to enteric viruses, pathogenic bacteria and protozoa.	Coliform organisms: <1 to 10^4/100 mL Other pathogens: controlled by treatment technology	< 1 to 2,000/mL
Nutrients			
Nitrogen	Nutrient source for irrigation; can also contribute to microbial growth.	10 to 30 mg/L	<1 to 30 mg/L
Phosphorus	Nutrient source for irrigation; can also contribute to microbial growth.	0.1 to 30 mg/L	<1 to 20 mg/L

Table 2.3. Comparison of criteria (maximum limits) for the irrigation of crops consumed by humans with reused wastewater by WHO, U.S. EPA, the State of California and some Mediterranean countries (national guidelines) (Angelakis et al., 1999).

Parameters	California[a] T-22 (1978)	U.S. EPA (1992)	WHO (1989)	Israel (1978)	Tunisia (1975)	Cyprus (1997)	France (1991)	Italy (1977)
Type of regulation	law	Guideline	Guideline	Law	law	Prov. Std.	guideline	Law
Minimum treatment required	advanced treatment	advanced treatment	stabilization ponds[b]	Secondary treatment[c]	stabilization ponds	tertiary treatment		secondary treatment
Total BOD_5 (mg/L)		10		15	30	10		
Dissolved BOD_5 (mg/L)				10				
SS (mg/L)		5[d]		15	30	10		
Turbidity (NTU)	2	2						
pH					6.5 to 8.5		as WHO	
Conductivity (dS/m)					7.0			
Dissolved O_2 (mg/L)	Present			0.5				
TC (MPN/100 mL)	2.2(50%)[e]	0[f]		2.2(50%); 12(80%)				

FC (MPN/100 mL)			1000			50	
Helminths (eggs/100 mL)[g]			1		<1	0	
Resid. Avail. Cl (mg/L)	Present	1.0		0.5			
Salinity							SAR<10[h]
Metals							Yes
Main treatment processes	Oxidation, clarification, filtration, disinfection	Filtration, disinfection	Stabilization ponds or equivalent	Long storage, disinfection	Stabilization ponds or equivalent	Filtration, disinfection	

[a]Spray irrigation. [b]Stabilization ponds in series with proper retention time. [c]Seasonal storage may constitute an equivalent to tertiary treatment. [d]If suspended solids are used instead of turbidity. [e]Not to exceed 23/100 mL in a single monthly test. [f]Not to exceed 14/100 mL at all times. [g]Nematodes such as Ascaris, Trichuris and hookworms. [h]$SAR=Na/((Ca+Mg)/2)^{1/2}$

2.2.1.2 Concerns of Water Reuse in Industry

Most of the industrial water reuse is for cooling water, which has the lowest water quality requirements compared with the industrial processes water. The cooling water represents the potential and possibility for municipal wastewater reuse. Five general water quality problems are encountered in this application.

Metallic Corrosion: the dissolution of metals by water.

Metallic corrosion is caused by the difference of electrical potential and is primarily a function of the pH and total dissolved solids (TDS), particularly chloride (Sunderberg et al., 1991). Contaminants such as TDS increase the electrical conductivity of the cooling water and thereby accelerate the corrosion rate. Chemical corrosion inhibitors can be added to control the corrosion (Metcalf & Eddy, 2003).

Scale: a kind of hard deposits, usually on the hot surface.

The cooling tower scaling problem is mainly caused by calcium scales (calcium carbonate, calcium sulfate, and calcium phosphate) and magnesium scales (magnesium carbonate and phosphate). These substances change from a soluble state in water to an insoluble state on the metal surface, which reduce the efficiency of heat exchange (Metcalf & Eddy, 2003). Some of factors that affect scaling also influence corrosion. Generally, cooling water that is scaling will tend to be non-corrosive and water that is non-scaling will tend to be corrosive. Indices relating calcium and alkalinity concentrations, pH, and TDS have been developed to predict the scaling

and corrosion tendency of the cooling water. The intention of treatment of cooling tower is to maintain a balance of slight scale forming tendencies, thus controlling corrosion while minimizing the impact on heat exchanger performance (Sunderberg et al., 1991).

Microorganisms: basic microorganisms such as aerobic bacteria, anaerobic corrosive bacteria, fungi, and algae (James, 1980).

The environment inside the cooling tower is ideal for the biological growth. Nutrients, particularly N and P, further encourage the growth of microorganisms, which may settle and bind other debris present in the cooling water. This problem needs more attention when reclaimed water is used for cooling water, because a greater concentration of organic matter exists in the reclaimed water.

Deposits: insoluble particulate matter in water.

The deposits consist of biological growths, suspended solids, silt, corrosion products, and inorganic scales. These insoluble particulate matters settle as a result of low flow velocity or adhere to hot or slime-covered surfaces and result in heat-insulating deposits. Chemical dispersants, chemical coagulation and filtration processes are effective in reducing the deposits.

Pathogens

The presence of pathogens (harmful viruses and bacteria) in the cooling water is a concern because of exposure of operators and maintenance personnel to the water and aerosols. Due to the

potential health risks, pathogens must be reduced to very low levels in order to make reclaimed water acceptable for unlimited contact (Sunderberg et al., 1991).

Wastewater reused in industry may mitigate the nutrient loadings in the final discharge, because of additional treatment applied at the power plant, oxidation and biological activity in the cooling system and the evaporation of ammonia in the cooling tower (Crook & Okun, 1987). A recent study showed that a cooling tower was used as a low-rate biofilm reactor for treating municipal wastewater and that there was high nitrogen and COD reduction in the system (Cloete *et al.*, 1999). Also, the use of reclaimed wastewater in cooling system is the least likely to produce public health hazards and more likely to receive public acceptance.

2.2.2 Application of UF and MF Membrane in Wastewater Reuse

To meet water quality objectives for water reuse applications, some physical, chemical and biological processes and operations have been developed to remove solids, organic matter, pathogens, metals, and nutrients. Compared with these technologies, membrane technology is becoming the more promising one with the high and consistent effluent water quality and economic competition. A significant potential for application of UF and MF for industrial and municipal wastewater reuse has been apparent (Christensen and Plaumann, 1981).

UF and MF membrane filtration process can effectively remove particles and microorganisms

from wastewater. This can be explained by the fact that the majority of the particles in the wastewater are larger than membrane pore size and the membrane acts as a selective barrier. Microfiltration membranes (0.1 μm) and ultrafiltration membranes (300,000 Dalton) can reject 100% of TSS and turbidity (Ahn and Song, 2000). Similar results have been observed with the different membrane pore and cut-off sizes (Alonso et al., 2001; Vera et al., 1998). Ahn and Song (2000) found the COD removal was about 80% with microfiltration (0.1 μm) and three different cut-off sizes of ultrafiltration membranes. Also, 47% and 61% COD reduction were found by other researchers (Alonso et al., 2001; Vera et al., 1998).

UF and MF membranes are excellent barriers for the total coliform, fecal coliform and fecal streptococci (Alonso et al., 2001; Vera et al., 1998). But the passage of small bacteria (0.2 to 0.3 μm sized) was observed although log reductions were >4 (Ghayeni et al., 1996).

Virus can be also removed by MF membrane owing to the deposited layer but not immediately after a backpulse (Peters and Pedersen, 1990). The 0.2 μm nominal pore size was challenged with polio virus (ca. 10^5/ mL). The results showed that virus removal was higher than 90% and retention was enhanced at lower pressure and in the presence of biomass/turbidity. The biomass/turbidity provided extra surface for adsorptive removal and formed a secondary layer on the membrane surface, which could increase the virus reduction (Ghayeni et al., 1996).

Microfiltration and ultrafiltration also provide a significant elimination of some metals: Fe, Zn, Al,

Ar, Cu and Mn (Alonso et al., 2001). The association of metals to suspended matter and to macromolecules in wastewater played a leading role in the effectiveness of filtration concentrating these elements.

Also, the effects of operating parameters on permeate quality were monitored. It was found that the changes in flux and suction mode had little effect on the values of TSS, turbidity and COD (Ahn and Song, 2000).

2.2.3 Application of NF and RO Membrane in Wastewater Reuse

In 1979, research showed that the reverse osmosis scheme could produce water at high and constant quality with an economically acceptable decline in membrane flux at a constant ratio of salt rejection (Hrubec *et al.*, 1979). Since then, RO membrane separation processes have been operated successfully for municipal wastewater reuse at several sites nationwide.

Through the results of five municipal wastewater reuse cases, it was summarized that RO was successful in the process for total dissolved solids reduction and viruses, pathogens, and bacteria removal (Filteau and Klinko, 1989). Water reclaimed from a multiple barriers system, including lime treatment, carbon adsorption, ozonation and reverse osmosis, exceeded the highest drinking water standards of the time (Rogers *et al.*, 1987).

Ernst and Jekel (1999) summarized their study conducted in Berlin, Germany, which investigated nanofiltration followed by ozonation for wastewater such that it could be used for groundwater recharge. The results show that nanofiltration effectively reduced dissolved organic C and adsorbable organic halides concentrations to less than 2 to 3 mg/L and less than 20 µg/L, respectively.

NF and RO membranes are applied for removal of dissolved ions and low molecular weight organic materials from liquids. When the wastewater is considered for direct portable reuse, these techniques are options.

2.3 Municipal Wastewater Plant Process and Water Quality

2.3.1 Conventional Secondary Treatment

Conventional secondary treatment is a biological process where the sewage after the preliminary and primary treatment, enters an aerobic reactor where organic matter in solution is used by microorganisms to grow (Gray, 1989). The reactor provides a suitable environment, oxygen and food for the microbial population to develop and use the organic matter in the wastewater. The incorporation of unsettleable material into the microbial biomass mixed liquor suspended solid (MLSS) leads to a much clearer wastewater. The last required operation is the separation of MLSS from the water by sedimentation. Secondary treatment can remove over 95% of the conventional pollutants.

There are two main types of biological processes used for the traditional secondary treatment: suspended growth and attached growth processes. In the latter, the micro-organisms are attached to an inert packing surface such as rock, gravel, slag, sand and a wide range of plastic and other synthetic materials. For the suspended growth processes, the micro-organisms mix freely within the wastewater and the biological cells which are separated after secondary settlement are returned to the reactor to maintain a high microbial density in order to achieve the maximum microbial breakdown of the wastewater. The secondary treatment phase may also comprise of other biological systems, both aerobic or anaerobic, or incorporated a mixture of several systems.

2.3.2 The Biological Nutrient Removal Process

Biological nutrient removal (BNR) can be described in general terms as the natural, biologically mediated reduction of nitrogen and/or phosphorus by the microorganisms found in activated sludge. By controlling the environmental conditions, one can provide the competitive growth environment for specialized bacteria responsible for nitrogen or phosphorus reduction over other bacteria. The removal of phosphorus by biological means is called as biological phosphorus removal and the biological removal of nitrogen is known as biological nitrogen removal.

2.3.2.1 Biological Phosphorus Reduction

Since the early 1980s, the success in full-scale plant biological phosphorus reduction has encouraged further use of this technology in wastewater treatment. The principle of biological

phosphorus reduction is that the phosphorus in the influent wastewater is incorporated into cell biomass, which subsequently is removed from the process as a result of sludge wasting. Compared with chemical technology using alum or iron salts to removal phosphorus, the principle advantages of biological phosphorus reduction are reduced chemical costs and less sludge production (Metcalf & Eddy, 2004).

A two-stage metabolic process is considered for biological phosphorus reduction (Figure 2.3). Under anaerobic conditions (in the absence of molecularly bound oxygen) heterotrophic phosphorus removing bacteria (Bio-P bacteria) release ortho-phosphorus into the mixed liquor to provide energy for the uptake and storage of short-chain volatile fatty acids (VFA). The most common VFA is acetic acid, which is the easiest VFA to be metabolized. Others include propionic and butyric acid. Energy and these readily available carbon sources are stored as intracellular polyhydroxybutyrate (PHB). When exposed to aerobic conditions, the stored PHB is metabolized to provide energy and carbon for new cell growth. Concurrent with the PHB oxidation the bacteria takeup ortho-phosphorus from the mixed liquor and store it as long polyphosphate molecules. It was (Chen et al., 2004) found that for a given long-term cultured biomass the more the soluble ortho-phosphate that was released in the anaerobic stage, the higher the amount of soluble ortho-phosphate that was taken up in the aerobic phase.

The conversion of phosphate from a soluble form (ortho-phosphate) to a polymeric molecule

stored within the cell biomass (polyphosphate), some phosphorus is ultimately removed from wastewater by the wasting of biomass from the bioreactor. Some phosphorus is recycled back to the anaerobic zone to provide energy for the uptake and storage of VFAs.

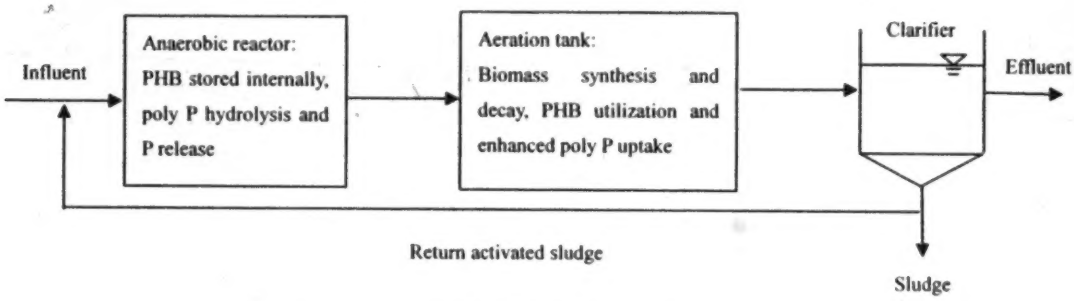

Figure 2.3. Biological phosphorus removal process.

Originally only that the genus Acinetobacter was believed to be responsible for bio-P reduction. Wagner *et at.* (1994) later found that this genus only plays a minor role in the removal of phosphorus from wastewater. Other organisms that were found in large populations in Bio-P removing activated sludge were Pserdomonas and Aeromonas bacteria. Some of these bacteria were found also have the capability of denitrification.

Some of the factors affecting biological phosphorus reduction include feed characteristics, temperature and operational conditions. Nitrate in the anaerobic phase inhibits the release of phosphorus (Yagci et al., 2003), because denitrifiers present in the anaerobic zone out-compete Bio-P bacteria for VFA, resulting in denitrification. An increase in the ratio of biochemical oxygen demand to total phosphorus in the influent enhanced phosphorus removal. An increase in pH was found to be a contributing factor to the increased release of phosphorus in the anaerobic

zone. Recently, laboratory studies have shown that phosphate removal was increased by up to 143% if the operational pH is held between the range of 5.5 to 6.5 for some isolated cultures from activate sludge samples. This is contrary to the belief that the optimal pH range for bio-P reduction was between 7.2 and 7.7. In accordance with the Arrhenius relationship, the biological phosphorus reduction reaction kinetic rates should decrease with temperature decrease. However, one study reported the efficiency of phosphorus reduction improved as the temperature was decreased in the range from 20 to 5 °C. The reason for the better system performance is apparently related to reduced competition for substrate in the anaerobic zones, which results in an increased population of phosphorus removing bacteria (Erdal *et al.*, 2003).

2.3.2.2 Biological Nitrogen Reduction (BNR)

The biological nitrogen removal process includes nitrification and denitrification. Nitrification is the term used to describe the two-step biological process in which ammonia (NH_4-N) is oxidized to nitrite (NO_2-N) and nitrite is oxidized to nitrate (NO_3-N). The biological reduction of nitrate to nitric oxide, nitrous oxide, and nitrogen gas is termed denitrification (Metcalf & Eddy, 2004).

The need for nitrogen removal arises from water quality concerns over (1) the effect of ammonia on receiving water with respect to DO concentration reduction and fish toxicity, (2) the need to provide nitrogen removal to control eutrophication, and (3) the need to provide nitrogen control for water-reuse applications including groundwater recharge. The typical effluent quality after biological nitrogen removal is 4 to 10 mg/L of total nitrogen based on a long-term average.

Soluble organic nitrogen and nitrate are the dominating compounds in the effluent (Henze, 1991).

2.3.2.2.1 Nitrification

The process of nitrification occurs under aerobic conditions and requires free oxygen molecules as a terminal electron acceptor. The oxidation of ammonium to nitrate is a two-step process: the first step is the oxidation of ammonium to nitrite; the second step is the oxidation if nitrite to nitrate (termed nitrification). Aerobic autotrophic bacteria are responsible for these processes in activated sludge and biofilm processes. *Nitrosomonas* and *Nitobacter* oxidize ammonia to nitrite and then to nitrate, respectively (Metcalf & Eddy, 2004).

Nitrosomonas bacteria:

$$2NH_4^+ + 3O_2 \longrightarrow 2NO_2^- + 2H_2O + 4H^+ \qquad \text{Equation 1}$$

Nitrobacter bacteria:

$$2NO_2^- + O_2 \longrightarrow 2NO_3^- \qquad \text{Equation 2}$$

Total oxidation reaction:

$$NH_4^+ + 2O_2 \longrightarrow NO_3^- + 2H^+ + H_2O \qquad \text{Equation 3}$$

Alkalinity required reaction:

$$NH_4^+ + 2HCO_3^- + 2O_2 \longrightarrow NO_3^- + 2CO_2 + 3H_2O \qquad \text{Equation 4}$$

Based on Equation 3, the oxygen required for complete oxidation of ammonia is 4.57 g O_2/g N. When synthesis of cell tissue is considered, the amount of oxygen required is less than 4.57 g

O_2/g N. Neglecting cell tissue, the amount of alkalinity required to carry out the reaction 4 is 7.14 g of alkalinity (as $CaCO_3$) or each gram of ammonia nitrogen (as N). Therefore, in biological nitrogen reduction, not only sufficient O_2 but also sufficient alkalinity is required. For locations with low-alkalinity waters, alkalinity is added at the wastewater treatment plant to maintain the nitrification reaction.

Nitrification is affected by a number of environmental factors including pH, temperature, toxicity, metals, and un-ionized ammonia. Under high ammonia concentration, the bicarbonate concentration is crucial for the nitrification process (Wett and Rauch, 2003). Optimal nitrification rates occur at pH values in the 7.5 to 8.0 range. The nitrification rate declines significantly at pH values below 6.8. In addition to low pH inhibiting nitrification, low temperature has a profound limiting effect on the growth of nitrifying bacteria. Metals, some inorganic and organic compounds are also of concern for nitrification process (Torrijos *et al.*, 2004).

2.3.2.2.2 Denitrification

Denitrification, an integral part of biological nitrogen removal, is the process by which nitrate is converted to molecular nitrogen with the absence of molecular oxygen. Compared to alternatives of ammonia stripping, breakpoint chlorination and ion exchange, biological nitrogen removal is generally more cost-effective and used more often.

Biological denitrificaton involves the biological oxidation of many organic substrates in

wastewater treatment using the oxygen in nitrate or nitrite as the electron acceptor instead of molecular oxygen. Therefore, the presence of nitrate or nitrite and the absence of molecular oxygen is necessary for denitrification. In biological nitrogen removal process, biodegradable organic matter, a product from endogenous decay and exogenous source (methanol or acetate) are the typical electron donors. Based on different electron donor, the denitrification reaction can be written (Metcalf & Eddy, 2004).

Biodegradable organic matter:

$$C_{10}H_{19}O_3N + 10\ NO_3^- \longrightarrow 5N_2 + 10CO_2 + 3H_2O + NH_3 + 10OH^- \qquad \text{Equation 5}$$

Methanol:

$$5CH_3OH + 6NO_3^- \longrightarrow 3N_2 + 5CO_2 + 7H_2O + 6OH^- \qquad \text{Equation 6}$$

Acetate:

$$5CH_3COOH + 8NO_3^- \longrightarrow 4N_2 + 10CO_2 + 6H_2O + 8OH^- \qquad \text{Equation 7}$$

Both heterotrophic and autotrophic bacteria have the ability of denitrification. The heterotrophic organisms include the following genera: *Achromobacter, Acinetobacter, Agrobacterium, Alcaligenes, Arthrobacter, Bacillus, Chromobacterium, Corynebacterium, Flavobacterium, Hypomicrobium, Moraxella, Neisseria, Pseudomonas* (Metcalf & Eddy, 2004). *Pseudomonas* species are the most common and widely distributed of all the denitrifiers, and have been shown to use a wide array of organic compounds. Autotrophic nitrifying bacteria, such as *Nitrosomonas europaea*, can use nitrite to oxidize ammonia and produce nitrogen gas, when dissolved oxygen is

not present (Metcalf & Eddy, 2004).

2.3.3 Membrane Bioreactor

The use of membranes as separation technology for biological treatment has created a new technique for the wastewater treatment. Typical membrane bioreactors (MBRs) are illustrated in Figure 2.4-a, the membrane separation unit is internal, immersed in the bioreactor. Treated effluent is withdrawn from the bioreactor with the application of a differential pressure. On Figure 2.4-b, the membrane separation unit is external to the bioreactor.

MBR system shows promise as a means of treating very high organic nitrogen wastewater without dilution. Using 0.2 μm membrane, rejection of 70 to 90% of residual COD was achieved even if the biological treatment process did not perform properly (Tay et al., 2003). With these advantages, MBR system has become more attractive in the field of municipal wastewater and industrial wastewater treatment for the past decade.

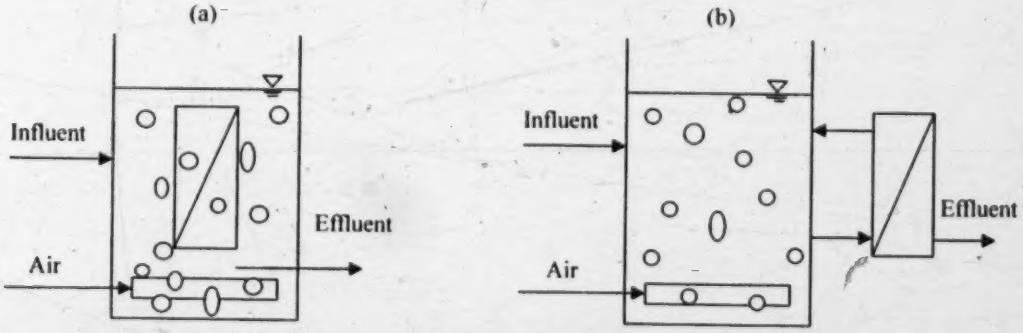

Figure 2.4. Schematic flow diagrams for the membrane bioreactor activated sludge process: (a) with internal membrane biosolids separation unit and (b) with external biosolids separation unit.

A 5 m^3/day capacity of MBR pilot plant was installed to treat the optic-electronic industrial wastewater. The COD, TOC and BOD_5 were reduced an average of 94%, 96% and 98%, respectively. Furthermore, the effluent did not contain suspended solids. Only a small concentration of ammonia nitrogen was found in the effluent. The effluent of TKN, NOx-N and COD can fall below 20 mg/L, 30 mg/L and 50 mg/L, respectively (Chen et al., 2003).

A 3.5 L membrane sequencing batch reactor was used for the treatment of a wastewater coming from the beam house section of a tannery. The system was operated for a period of 150 days, with no sludge removal during the whole period of operation. Removal efficiencies close to 100% in ammonium and 90% in COD were achieved and the TKN removal efficiency ranged from 60 to 90% (Goltara et al., 2003). A pilot-scale (10 m^3/d) anoxic/oxic MBR (A/O MBR) was tested for dyeing wastewater treatment of woolen mill without wasting sludge in 125 days operation. Results showed that the removal rates of COD, BOD_5, colour, and turbidity were 92%, 98%, 74% and 99%, respectively (Zheng et al., 2003).

To enhance biological phosphorus removal (EBPR), an innovative process sequencing anoxic/anaerobic membrane (SAM) bioreactor was developed. By supplying strict anaerobic conditions without an internal recycle, the SAM system can yield 93% phosphorus reduction efficiency. The nitrogen removal efficiency of the SAM was about 60%, which was slightly lower than that of the modified Luzack-Ettinger (MLE) type MBR process, in which the mixed liquor

was recycled continuously from aerobic zone to anoxic zone. However, it should be noted that the hydraulic retention time of the SAM process in the anoxic condition was 2.3 times shorter than that of MLE-type MBR process (Ahn et al., 2003).

2.4 Gold Bar Wastewater Treatment Plant

The Gold Bar Wastewater Treatment Plant (GBWWTP), located on the south bank of the North Saskatchewan River in Edmonton, Alberta, consists of pre-treatment, primary treatment, activated sludge secondary treatment, medium-pressure UV microorganism reduction processes. The primary treatment capacity is 910 million litres per day (ML/day) and the secondary treatment capacity is on an average 310 ML/day of wastewater, with a peak capacity of 420 ML/day.

In 2002, the secondary treatment was converted to full biological nutrient removal treatment. The GBWWTP BNR process (Figure 2.5) includes pre-anoxic, anaerobic, anoxic and aerobic zones. GBWWTP produces excellent quality effluent on a consistent basis (Based on 2001): BOD_5 5.7 mg/L, TSS 9.5 mg/L, TP 1.0 mg/L, NH_3-N 9.5 mg/L (winter) 6.4 mg/L (summer), FC 31 CFU/100 mL (Heise, 2002).

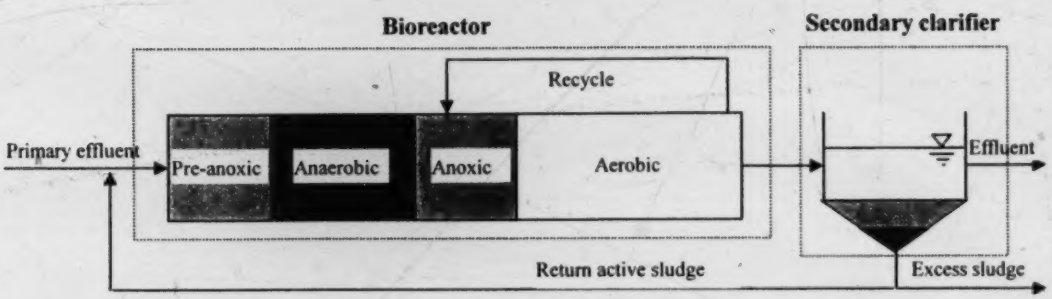

Figure 2.5. GBWWTP full-scale BNR process.

3. Methods and Materials

3.1 Experimental Equipment and Control

3.1.1 Membrane System

The experimental set-up, a classical UF filtration pilot plant (Figure 3.1), was located at the GBWWTP, Edmonton, Alberta. This unit was supplied with two skids, a membrane tank provided by GBWWTP and a process equipment skid provided by ZENON Environmental Inc. The schematic of this pilot plant is shown in Figure 3.2 and the characteristics of this pilot plant are listed in Table 3.1.

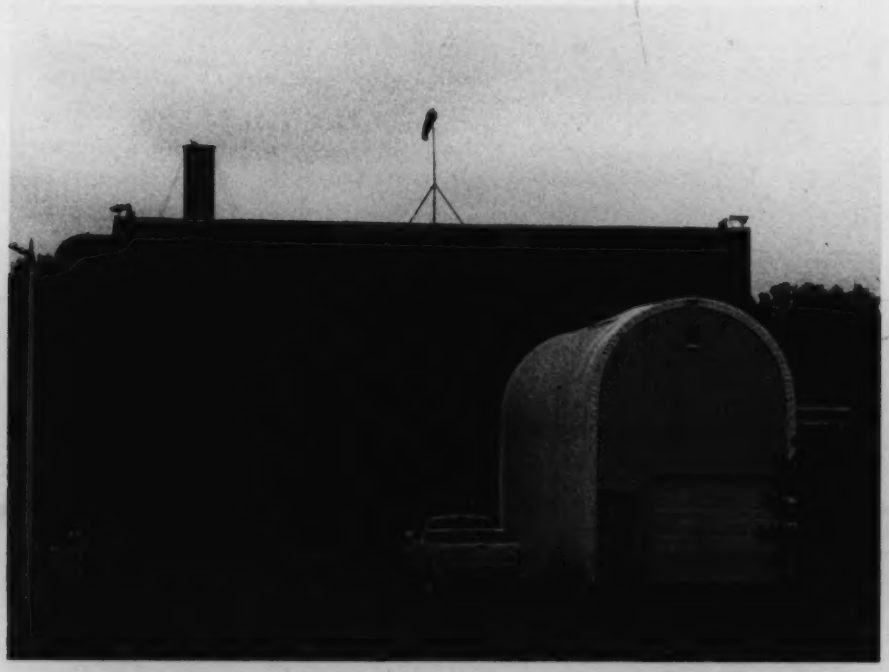

Figure 3.1. ZeeWeed 500D membrane pilot plant in GBWWTP.

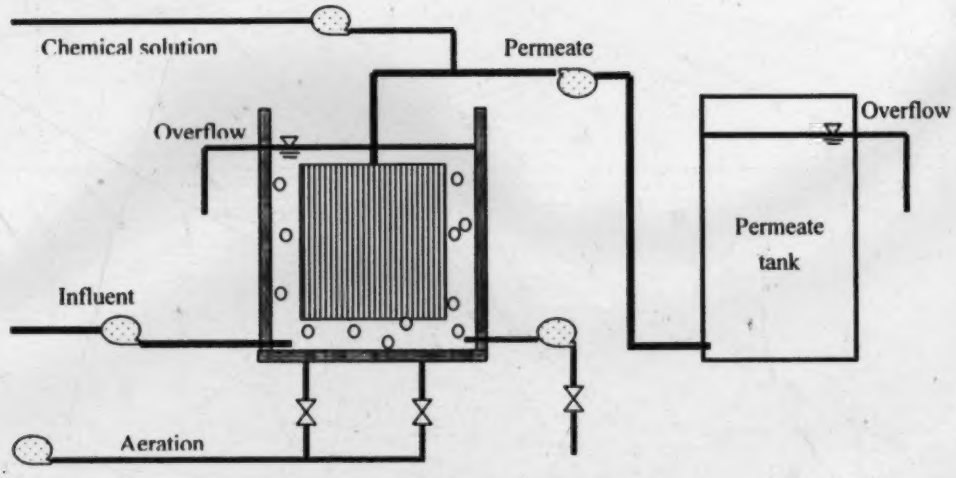

Figure 3.2. Experimental set-ups.

The operation of this unit was automated using a Programmable Logic Controller (PLC) that monitored and adjusted most of the vital functions. The basic control procedure was presented in Appendix A. The Human Machine Interface (HMI) (Figure 3.3) coupled with the PLC provided an interface between PLC and the operator. The operator was able to adjust set points or start and stop the unit through HMI. In case of a particular problem, the PLC provided alarms to draw the operator's attention.

Before the raw water entered the membrane tank, it passed through an in-line strainer (3 mm) to remove the large solid contamination and protect the membrane. When the pressure of the strainer built up, a manual cleaning was required.

In the membrane tank, the membrane tubes were aerated via the membrane aeration blower to remove solids which accumulated on the membrane surface. Air was distributed along the bottom of the membrane by an aeration tube, which was an integral component of the membrane system. The aeration also kept the membrane tank mixed and shook the membrane tubes to prevent the accumulation of solids on the tubes.

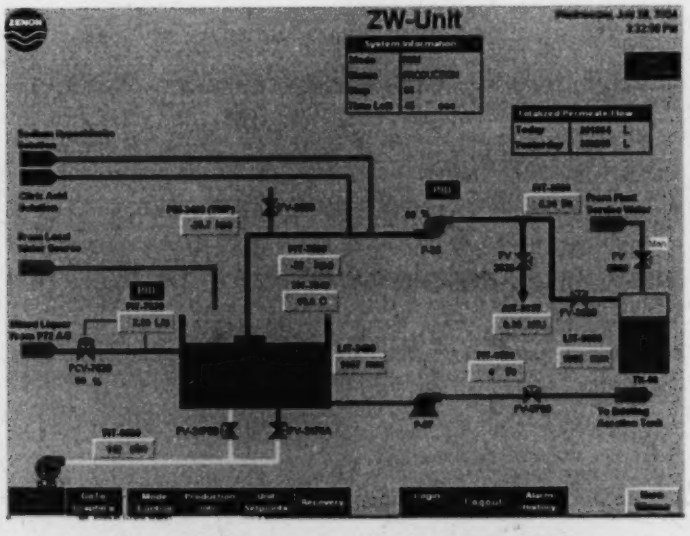

Figure3.3. Human machine interface (HMI).

A slight negative pressure was used to extract permeate through the membrane. During this process, solids were held back and accumulate on the membrane surface, which restricted the flow through the membrane and caused a higher negative pressure to be applied. Periodically reversing the permeate flow through the membranes termed "backpulse" was used to detach

solids from the membrane surface. After the unit operated continuously for an extended period of time or the transmembrane pressure (TMP) increased to above a preset value, chemical cleaning was required to clean the membrane, increase the high permeate flux, and lower the TMP.

Table 3.1. Characteristics of the UF unit.

Type of Membrane	ZeeWeed 500D
Nominal Pore Size	0.04 μm
Total Membrane Area	758.4 m^2
Flow Type	Outside-in
Highest Transmembrane Pressure	50 kPa
Average Daily Flow Rate	284 m^3
Maximum Daily Flow Rate	425 m^3
Maximum Backpulse Flow Rate	8.3 L/s
Chemical Cleaning	Citric Acid & Sodium Hypocloride

3.1.2 ZeeWeed Ultrafiltration Membrane

Zenon Environmental Inc. supplies two ultrafiltration membrane systems, ZeeWeed 500 series and ZeeWeed 1000 series. Besides the different configuration, the ZeeWeed 1000 series membranes are targeted for treating feed waters containing low amounts of suspended solids, while ZeeWeed 500 series membranes are intended for applications with medium to high-suspended solids concentrations of feed water (ZENON, 2004). For ZeeWeed 500 series, normally a cassette is the smallest operable unit of the filtration system and 4 to 64 modules are included in one cassette. The following Figure 3.4 shows the schematic of ZeeWeed 500 module.

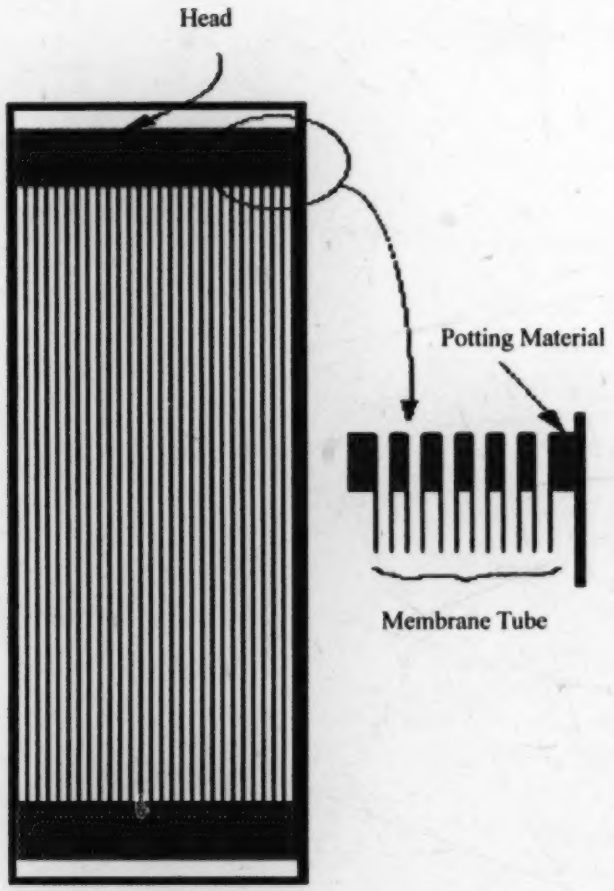

Figure 3.4. The schematic of ZeeWeed 500 module.

In this project, the ZeeWeed 500D cassette (Figure 3.5) was used as tertiary treatment process at GBWWTP. This cassette had space for 48 modules, but only 24 modules were installed. The membrane surface area for each module is 31.6 m^2, so the total membrane surface area of this cassette is 758.4 m^2.

Figure 3.5. Membrane cassette used for this project.

Each ZeeWeed module typically consists of hundreds of tiny hollow fibers bundled together and sealed at each end with a resin to form a watertight seal between the permeate and feed streams. With this type of configuration, hollow fiber modules have a very high packing density. This provides a large surface area for permeate production.

3.1.3 Production Process

One ZeeWeed membrane operating cycle included two processes: production and backpulse. During the production, a small vacuum was created within membrane fibers by the process pump and water was drawn from the membrane tank through the pores in the membrane fibers.

Meantime, solids larger than the membrane pores were rejected. The schematic of this process is shown in Figure 3.6. After a period of time, particles deposited and accumulated on the membrane surface and formed a solid layer even though the aeration outside the fibres somewhat reduced the deposition rate. Then, the backpulse was applied to detach this solid layer and improve membrane performance. In this system, the backpulse was an on-line self-cleaning process. During the backpulse, the process pump, which was a reversible pump, pumped permeate water through the membrane surface to "blow" solids off the surface (Figure 3.7). The frequency, duration and intensity of backpulse were defined by the operator with respect of feed water quality and flux.

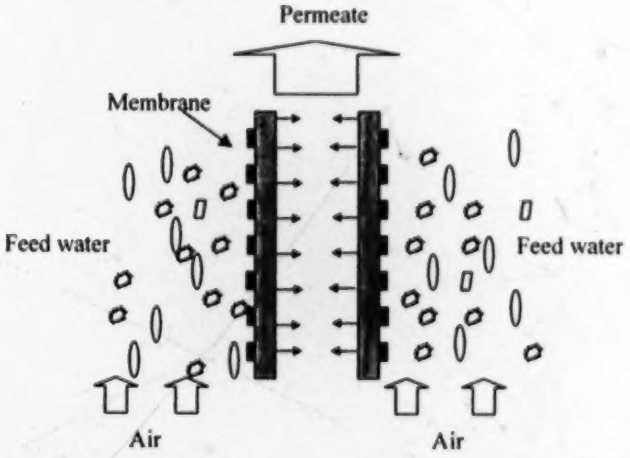

Figure 3.6. Production process.

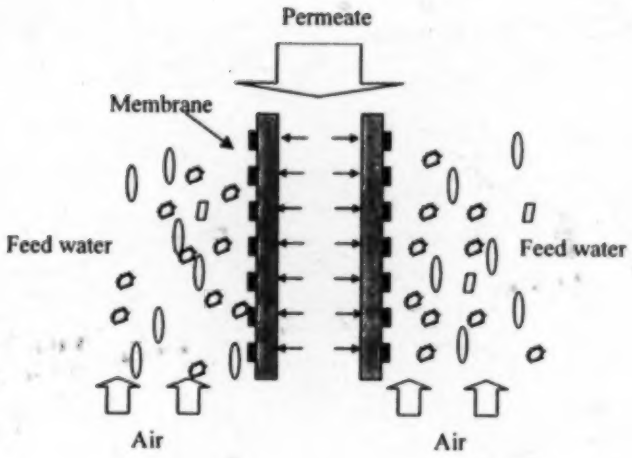

Figure 3.7. Backpulse process.

3.1.4 Chemical Cleaning

To maintain and recover membrane performance, maintenance and recovery cleaning were employed on this ZeeWeed® membrane system. By comparison of these two types of cleaning, the maintenance cleaning required less chemical and took shorter time, however the cleaning effectiveness of maintenance cleaning was lower than the recovery cleaning. Both of these procedures were automated after the switch from normal operation to cleaning operation, but the operator was required during the cleaning process.

Depending on the type of fouling on the membranes, either a sodium hypochloride solution or citric acid solution was used. The concentration, amounts and design pH of the cleaning chemical solutions are given below (Table 3.2). In practice, these requirements were not met exactly.

Table 3.2. Chemicals required for cleaning the ZeeWeed system.

Cleaning chemical	Purpose	Design wash concentration	Approx. flow	Design pH
NaOCl at 12% for full tank maintenance clean	Organic cleaner	250 mg/L	0.5 L/min	
Citric acid at 50% for full tank maintenance clean	Scale removal	2,000 mg/L	0.85 L/min	
NaOCl at 12% for empty tank maintenance clean	Organic cleaner	500 mg/L	1.7 L/min	
Citric acid at 50% for empty tank maintenance clean	Scale removal	2,000 mg/L	1.4 L/min	
NaOCl at 12% for recovery clean	Organic cleaner	2,000 mg/L	6.81 L/min	Max. 10
Citric acid at 50% for recovery clean	Scale removal	8,000 mg/L	68 L/min	2.5 to 3.5
Neutralization chemicals	**Purpose**		**Approx. quantity**	
Sodium Hydroxide	Adjust of pH		65 L/clean	
Sodium bisulfite	Dechlorination		51.2 L/clean	

3.1.4.1 Maintenance Cleaning

For maintenance cleaning, there are two subtypes "full tank" and "empty tank". Empty tank maintenance cleaning requires that the feed to the membrane tank is stopped and the membrane tank is completely drained. Over a 90-minute period, a certain amount of chemical is backpulsed through the membrane in regular pulses followed with permeate water flush. As the cleaning chemicals remain in direct contact with the membrane and are not diluted by the water in the membrane tank, this method of cleaning is more effective than full tank maintenance cleaning.

For the full tank maintenance cleaning, approximately 250 mg/L of sodium hypochlorite (or 2 g/L citric acid) is backpulsed through the membrane in regular pulses over a one-hour period. During the full tank maintenance cleaning, the feed to the membrane tank is stopped, and the membrane

aeration blower is also stopped. However, intermittent aeration has to be applied to the membrane tank in order to wash chemical and solids away from the membrane fibers. The maintenance cleaning procedure is listed in Appendix B.

3.1.4.2 Recovery Cleaning

A recovery cleaning is required to restore the permeability of the membrane, when the automatic backpulse and maintenance cleaning cannot maintain membrane performance above the specified level, or when permeability declines to less than 50% of the initial stable permeability or below $2.7 * 10^{-7}$ $m^3/m^2/s/kPa$. This will generally occur when the TMP exceeds 35 to 42 kPa under peak flow conditions. If recovery cleaning is not executed at this time, further fouling may become irreversible. A brief summary of the steps for a recovery cleaning is listed as follows:

1. Note and record TMP and flow readings for one complete permeation cycle at intervals of 2 minutes each.
2. Switch to the mode of operation screen.
3. Switch the mode from "Run" to "Wash".
4. Stop influent pump.
5. The ZeeWeed system remains in soak until the operator selects "Chlorine R Clean" or "Acid R Clean".
6. Under the control of the PLC, the membrane tank is drained.
7. The membranes are automatically backpulsed with a chemical solution.
8. The permeate/backpulse pump is stopped by PLC automatically.

9. The HMI displays the banner "Tank Filling Complete". The operator fills the membrane and then touches the "Tank Filling Complete" button.

10. The membranes will begin to be soaked for 24 hours.

11. The HMI displays the banner "Confirm Neutralization". The operator manually neutralizes the cleaning solution in the membrane tank and no equipment is supplied for this operation. The cleaning solution can be neutralized with sodium hydroxide for pH adjustment after an Acid recovery cleaning and sodium bisulfite for chlorine removal. Chlorine removal decreases pH and also requires sodium hydroxide for pH adjustment.

12. The operator presses "Confirm Neutralization" button on the HMI.

13. The PLC controls the recovery cleaning till the operator restart the feed pump to refill the membrane tank.

14. Switch the mode from "Wash" to "Run" to return to production.

For the detailed information of recovery cleaning steps by the PLC, please refer to Appendix C.

3.1.5 Parameters for Membrane Performance

Many parameters are used to characterize the membrane performance, such as transmembrane pressure, recovery, flux, permeability and log removal. The first three of those parameters are related to the membrane operation and the log removal is related to the permeate water quality.

Transmembrane Pressure

Transmembrane pressure (TMP) is defined as the difference between the pressure on the outside

of the membrane (hydrostatic pressure) and the applied vacuum in the inside of the membrane created by the process pump. For the ZeeWeed 500D membrane system, the TMP should not be higher than 65.5 kPa during production. The location where TMP is calculated is shown in Figure 3.8.

Pressure Inside

= Measured pressure – pressure due to elevation of pressure transmitter

= PIT 3523 + (A/C) Equation 8

Pressure Outside

= Pressure due to water above membranes

= (Membrane tank transmitter level – distance from level transmitter to top of membrane) / C

= (LIT 3426 – B) / C Equation 9

TMP = Pressure outside – Pressure inside

= ((LIT 3426 – B) / C) – (PIT 3523 + (A/C)) Equation 10

PIT 3523: Measured pressure

LIT 3426: Membrane tank transmitter level

A: Pressure due to elevation of pressure transmitter

B: Distance from level transmitter to top of membrane

C: Elevation to pressure conversion factor

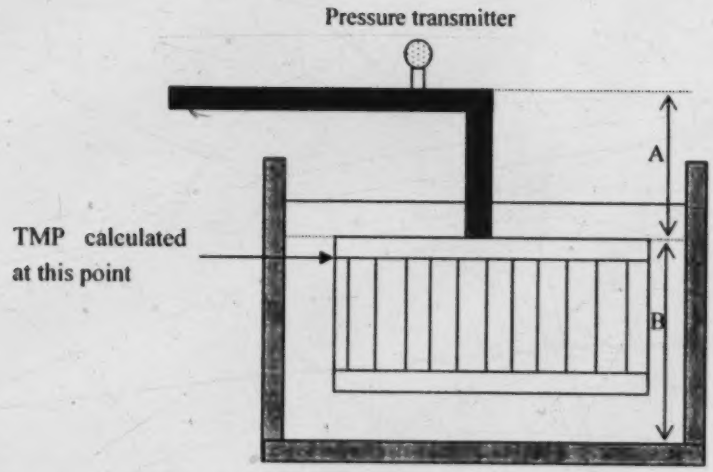

Figure 3.8. TMP calculated point.

Highest Transmembrane Pressure (HTMP)

In this project, highest transmembrane pressure (HTMP) in one operating cycle was another very useful parameter. HTMP was usually the TMP just prior to the backpulse in one operating cycle. Defauted in the PLC control system, the permeate pump run by 35% of a designed rate at the beginning of one operating cycle. After 3 seconds, the pump was released to flow control and TMP control. In one operating cycle, the system compensated the permeate water used for the last backpulse. The PLC system calculated what flow rate it should reach in this cycle to reach the net permeate flow set point for the period of this operating cycle. After the system reached the flow, it maintained that flow, but the TMP was still increasing to keep the flow. Sometimes the system could not reach the desired flow due to some other limitation, but it kept trying till the system highest TMP (50 kPa). So the highest TMP in one operating cycle was always at the end of the

cycle. Figure 3.9 and Figure 3.10 present the sample of the variation of TMP and permeate flux in operating cycles.

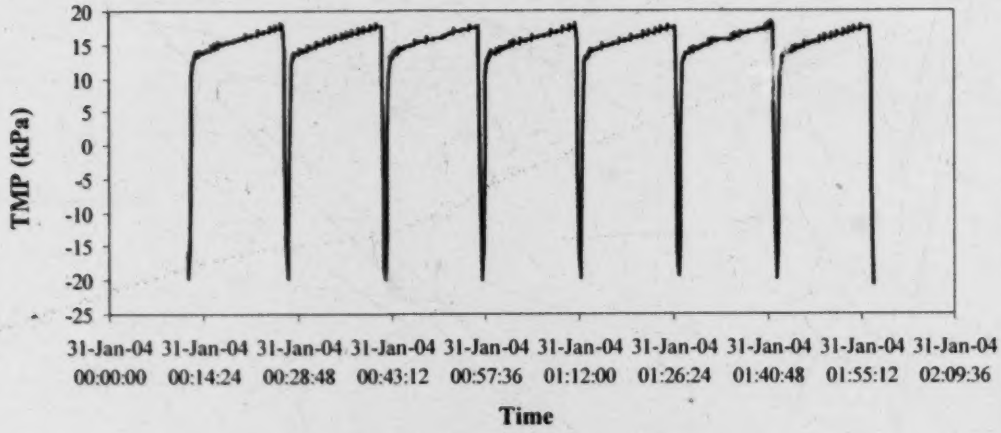

Figure 3.9. TMP versus time in operating cycles.

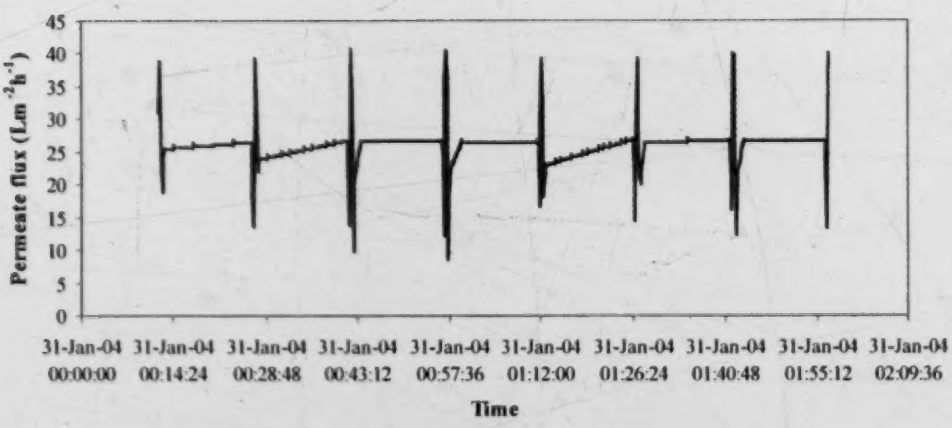

Figure 3.10. Permeate flux versus time in operating cycles.

In the production process, solids were concentrated in the membrane tank, so a portion of the

water in the membrane tank has to be wasted by overflow to avoid the continuous increase of the solids concentration in the membrane tank. So the water could not be recovered at 100%. Therefore, the definition of recovery rate was very crucial to evaluate the membrane performance in wastewater treatment processes. The recovery rate can be simply defined as the ratio of permeate flow rate to feed flow rate.

Recovery = Permeate flow rate / Feed flow rate

Flux

Flux is defined as the ratio of flow rate to membrane surface area (Equation 11) and used to indicate how hard the membranes are working. The flux range of ZeeWeed 500 for drinking water applications is from 50 to 75 L / m^2·hr; for wastewater applications, the flux range is from 15 to 35 L / m^2·hr.

Because temperature significantly affects membrane performance due to variations in water viscosity, temperature correction is necessary to ensure that permeability trends are not the result of temperature variations.

Flux = Flow rate/ Membrane area Equation 11

Flux @ T2 = ((Flux @ T1) * (Viscosity @ T1)) / (Viscosity @ T2) Equation 12

Karimi et al. (1999) provided the following relationship between the viscosity and temperature.

(For $T \leq 20$ °C) $\mu_T = e^{-0.0282 (T-20)}$ Equation 13

(For $T > 20$ °C) $\mu_T = e^{-0.021 (T-20)}$ Equation 14

Permeability

Permeability is the ratio of flux to TMP and its unit is L / m^2 / hr / kPa. Permeability is used to indicate how much energy is required to make permeate and normalize membrane condition (degree of fouling). The application of permeability will be used to indicate the membrane fouling condition in the next chapter.

Permeability = Flux / TMP = Flow / Membrane area / TMP Equation 15

Log reduction

Log reduction indicates the degree of removal or inactivation of pathogenic organisms through physical-chemical treatment of water. In this project, log removal is used to show the removal effectiveness of ZeeWeed 500D membrane for coliform microorganisms.

Log Removal = - log (Permeate concentration / Feed concentration) Equation 16

3.2 Test Runs

3.2.1 Bubble Test

Bubble testing is required to ensure that no damage has occurred to the fibers during shipping or installation and to test the permeate pipe connections that are completed during installation. When the new membrane was installed, the bubble test could not be operated until at least 24-hour operation. It was also necessary to conduct bubble test at various times during the production period as a maintenance procedure. In this project, at the beginning when the installation was just finished, in the middle of this project and at the end of this project, a total of three bubble tests

were conducted on the unit and no leakage was found.

Before a bubble test was conducted, the membrane tank must be filled with utility water; the system was completely shut down and the membrane was disconnected from the permeate pump. Then a pressure regulator was connected between the membrane and pressured air. The valve of the pressure regulator was slowly opened and the increase in the pressure reading on the pressure indicator was observed until the pressure increased to 37 kPa. The valve then was closed and the pressure drop was monitored. The membrane was carefully examined for air bubbles. The frequency and type of the bubbles determine what action, if any, must be taken. Bubbles are generally classified into four types (ZENON, 2003).

Type 1 bubbles appear as steady streams of large bubbles, typically 10 mm diameter or greater. If the stream appears to come from one of the permeate headers, it is likely caused by an improperly installed O-ring. Otherwise, it is likely a broken fiber.

Type 2 bubbles are characterized by steady streams of moderate size >1 mm and < 3 mm bubbles. This category covers problems including damaged fibers or "pin holes" in the membrane. Bubbles of this type indicate a minor leak that may or may not need repair. If larger leaks are found in the cassette or sub-group of cassette, these should be repaired and then a PDT performed to determine whether these smaller leaks need repair.

Type 3 bubbles appear as an intermittent stream of small bubbles <1 mm. This type of bubble is typically due to air passing through unwetted pores. Unwetted pores allow passage of air but not water and do not need to be repaired.

Type 4 bubbles are those that gather under the top header of a ZeeWeed 500 element and release when sufficient volume has collected. They come to the surface in one moderate to large bubble. Typically, type 2 or type 3 bubbles gather to form this type of bubble. Investigate by slowly raising the cassette to determine the cause of the bubble.

3.2.2 Tracer Test

To find out the hydraulic regime of the membrane tank, two tracer tests were conducted on January 9 and January 13, 2004. Sodium chloride was used as a tracer agent and introduced to the membrane tank as an impulse input. The concentration of sodium ion was monitored for both permeate and overflow. The operation parameters during the tracer test are listed in Table 3.3.

Table 3.3. Operating parameters for the tracer test.

	January 9	January 13
Influent flow rate (L/s)	3.17	3.08
Permeate flow rate (L/s)	2.47	2.71
Overflow flow rate (L/s)	0.7	0.36
Operating time (min)	15	15
Duration of backpulse (s)	30	30
Blower (m^3h^{-1})	247	240
NaCl added (g)	1600	1614

Figure 3.11, and 3.12 show the change of sodium ion concentration with the time for permeate

and overflow. It was found that there was no significant difference for sodium ion concentration in the permeate flow and overflows. Therefore, the influent flow rate and membrane tank volume were used to calculate the theoretical hydraulic retention time in the equation $\Gamma=Q/V$. For this tracer test, it took several minutes to dose sodium chloride to the membrane tank, so there was a short lag of the peak point. By ignoring this, the membrane tank was modelled as a continuous flow stirred tank (CFST) and the equation $Y=C_o e^{-t/\Gamma}$ was used to calculate the hydraulic retention time for the membrane tank. The calculation results are listed in Table 3.4.

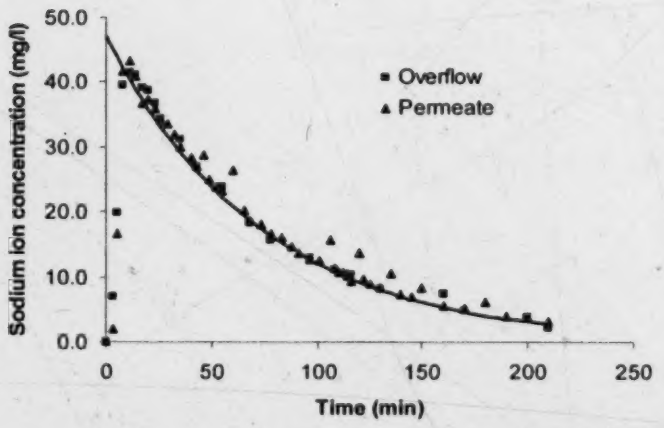

Figure 3.11. The sodium ion concentration versus time on January 9.

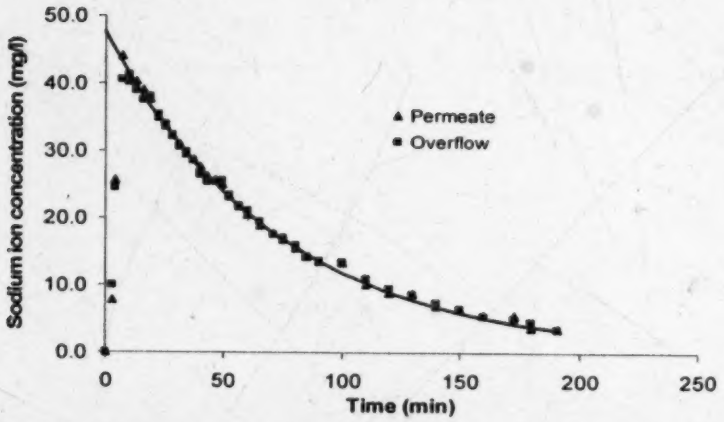

Figure 3.12. The sodium ion concentration versus time on January 13.

Table 3.4. Measured and theoretical hydraulic retention time for pilot membrane plant.

Date	Retention Time (min) of Permeate		Retention Time (min) of overflow	
	Measured Value $Y=C_oe^{-t/T}$	Theoretical Value $\Gamma=Q/V$	Measured Value $Y=C_oe^{-t/T}$	Theoretical Value $\Gamma=Q/V$
Jan 9	82.0	70.1	73.5	70.1
Jan 13	71.9	72.1	76.3	72.1

3.3 Experiment Design

To optimize the operating parameters, different operating parameter set points have been examined for two different feed waters: FE after UV and FE before UV (Table 3.5 and 3.6). Every time before new operating parameter set points were changed, NaOCl maintenance cleaning was conducted to recover the membrane hydraulic flux and make each experiment consistent.

The cleaning effectiveness of both maintenance cleaning and recovery cleaning was also

examined. To check the cleaning effectiveness of both maintenance and recovery cleaning, the operating parameter set points before and after the cleaning were kept same and the comparison of the operating parameters before and after cleaning demonstrated the effectiveness of cleaning. All the operating changes made during this experiment are listed in Appendix D.

3.4 Sampling Design

Samples were collected three times per week from January 19, 2004 to September 13, 2004. Every time three grab water samples (influent, permeate and overflow) were collected and split into sub-samples for different analyses. Total suspended solid (TSS), 5-day biochemical oxygen demand (BOD_5), chemical oxygen demand (COD), total Kjeldahl nitrogen (TKN), ammonia (NH_3H), total oxidized nitrogen (TON), total phosphorus (TP), total hardness (TH), total organic carbon (TOC), total alkalinity (TA), turbidity, pH, particle count, calcium, magnesium and chloride were analyzed for the permeate. TSS, NH_3H, TON, TP and pH were analyzed for influent water sample. Occasionally, silt density index (SDI) and coliphage were analyzed for permeate, influent and overflow. 200 mL influent and permeate samples were collected separately each time for the total coliform (TC) and fecal coliform (FC) test. For the overflow, only TSS, TC and FC were analyzed. In certain time, metal analyses were taken for permeate, influent and overflow.

Table 3.5. Summary of the operating parameters set points for FE after UV used as feed.

Operating time (min)	Flux [A] (L/m^2 h)	BP (s)	duration	Aeration (m^3/h)	Water recovery (%)	Date (D/M)
15	23.7	30		237	70%[B]	19/1 to 19/2
						21/4 to 22/4
						20/5 to 28/5
15	28.4	30		237	70%[B]	30/3 to 02/4
15	19.0	15		237	70%[B]	12/3 to 16/3
15	23.7	30		203	70%[B]	26/3 to 30/3
15	23.7	15		237	70%[B]	09/3 to 12/3
						22/3 to 26/3
10	23.7	30		237	70%[B]	16/3 to 22/3
						16/4 to 20/4
10	23.7	15		237	70%[B]	19/2 to 26/2
10	28.4	30		203	70%[B]	08/4 to 16/4
10	28.4	30		237	70%[B]	02/4 to 08/4
25	23.7	30		237	70%[B]	26/2 to 01/3
						14/5 to 20/5
25	19.0	30		237	70%[B]	09/3 to 11/3
25	14.3	30		237	70%[B]	01/3 to 09/3
25	14.3	30		203	70%[B]	04/5 to 14/5
10	25.1	24		237	70%[B]	26/4 to 30/4
15	19.0	15		237	70%[B]	
15	28.4	15		237	70%[B]	20/4 to 21/4
12.5	26.1	23		237	70%[B]	22/4 to 26/4
10	25.1	24		237	70%[B]	
10	28.4	15		203	70%[B]	30/4 to 04/5

[A] flux means net flux set point, not the actual operating one.

[B] water recovery is roughly calculated based on the influent and effluent flow rate.

Table 3.6. Summary of the operating parameters set points for FE before UV used as feed.

Operating time (min)	Flux[A] (L/m^2 h)	BP duration (s)	Aeration (m^3/h)	Water recovery (%)	Date (D/M)
15	23.7	30	237	70%[B]	11/6 to 28/6 14/7 to 30/7
15	23.7	40	237	70%[B]	28/6 to 14/7
25	14.3	30	237	70%[B]	28/5 to 02/6
25	23.7	30	237	70%[B]	02/6 to 07/6
15[D]	23.7	30	237	90%	30/7 to 24/8
15[D]	23.7	30	237	95%	24/8 to 31/8
10	26.1	30	237	70%[B]	07/6 to 11/6

[A] flux means net flux set point, not the actual operating one.

[B] water recovery is roughly calculated based on the influent and effluent flow rate.

3.5 Analytical Methods

All analytical methods were based on methods outlined in *Standard Methods for the Examination of Water and Wastewater, 19th Edition* (APHA, et al., 1995) and the Alberta Environment Center's (AEC) Methods Manual for Chemical Analysis of Water and Wastes (Dieken et al., 1996), except that Silt Density Index (SDI) was based on American Society For Testing Materials (ASTM D 4189-95) standard. Quality control samples (standards, blanks and spikes) were included within analytical sets to ensure quality of data. Sample data validation and acceptance was subject to the quality control guidelines and limits of GBWWTP laboratory.

3.5.1 TSS Determination

Total suspended solid (TSS) was determined with *Standard Methods* (APHA, et al., 1995) 2540 D. The water sample was filtered through a weighted 25 mm fiberglass filter and a crucible; the residue retained on the filter was dried to a constant weight at 103 to 105 °C. The gain of the

weight was the total suspended solids. As TSS in permeate was expected to be low, a larger volume of sample was required to pass through the filter. The detection limit was 0.6 mg/L.

3.5.2 BOD_5 Determination

The BOD_5 was analyzed based on *Standard Methods* (APHA, et al., 1995) 5210 B. A 256 mL diluted sample was incubated at $20 \pm 1^{\circ}C$ for 120 ± 2 hours in a temperature controlled stainless steel water bath; BOD dilution water was prepared according to *Standard Methods*. Dissolved oxygen was measured before and after the incubation. The difference of dissolved oxygen was the BOD_5 value. The detection limit for BOD_5 value was 2 mg/L.

3.5.3 COD Determination

COD was determined with *Standard Methods* (APHA, et al., 1995) 5220 D based upon the closed reflux, colorimetric HACH method. The detection limit was 2 mg/L for low reference (COD $\leqslant$ 150 mg/L) and 5 mg/L for high reference range (COD > 150 mg/L).

3.5.4 Ammonia Determination

The GBWWTP laboratory utilized the automated phenate colorimetric method based upon the AEC method 219 (Dieken et al., 1996) for ammonia determination. The detection limit was 0.013 mg/L.

3.5.5 Total Oxidized Nitrogen Determination

The GBWWTP laboratory utilized AEC method 2359 (Dieken et al., 1996) for the determination of total dissolved NO_3^- and NO_2^- in wastewater. The detection limit was 0.006 mg/L.

3.5.6 Total Kjeldahl Nitrogen Determination

The semi-automated block digestion, phenate colorimetric method was used to determine the total Kjeldahl nitrogen in the water sample. This method was referred to AEC 235 (Dieken et al., 1996).

3.5.7 Total Phosphorus Determination

The semi-automated block digestion, ascorbic acid, molybdenum blue colorimetric method was used to determine the total phosphorus content in the water sample. This method was referred to AEC 582 (Dieken et al., 1996).

3.5.8 Turbidity

Turbidity was determined with the nephelometric HACH method 2130 B using a HACH 2100AN turbidimeter.

3.5.9 pH

An Acumet 950 pH meter was used to determine the mixed liquor pH. The meter was calibrated with a three-point calibration using pH buffers 4, 7 and 10.

3.5.10 Coliforms

The GBWWTP utilized Standard *Methods* 9222 B and 9222 D (APHA, et al., 1995) to determine total coliforms and fecal coliforms respectively. This method was based upon the membrane filtration method using MENDO agar and FC agar.

3.5.11 Coliphage

A modified Double Agar Layer (DAL) method, as suggested by Grabow (2001) was used for coliphage enumeration. *E.coli* C. was used as the host culture. A brief description of the DAL method is provided below.

1. Bottom agar was first prepared by following procedure. Dissolve 30 g of Trytocase soy broth (TSB) and 15 g of agar in 1000 mL of laboratory-grade water. Autoclave the aqueous soution at 121°C for 30 minutes. After cooling down in a water bath at 47 °C and the solution was aseptically poured into Petri plates. After solidifying the bottom agar was positioned up down for later use.

2. A fresh overnight (10 to 12 hours) culture of the host bacteria was prepared using TSB broth.

3. The top agar using 30 g of TSB and 7 g of agar in 1000 mL of laboratory-grade water was prepared and autoclaved in an Erlenmeyer flask and keep in a water bath at 47 °C.

4. 3 mL of the host culture was added in 100 mL of top agar, vortex mixes, and let it stand in the water bath for 3 minutes.

5. 5 to 100 mL of the sample (depending on the expected coliphage concentrations) was added to

the soft agar and host culture, mixed gently and kept in the water bath for 3 minutes.

6. Approximately 20 mL of the soft agar was poured onto each 150 mm Petri plate and 7 mL onto each small Petri plate containing the bottom agar. After solidifying, it was incubated inverted at 37°C ± 0.1°C for 8 to 12 hours.

3.5.12 Metals Determination

Standard Methods 3120 B Inductively Coupled Plasma (ICP) method (APHA, et al., 1995) was used to determine the metal concentration in water samples.

3.5.13 Particle Count

L & H Environmental water particle sensor was used to provide particle count data over the particle size range from 2 to 200 microns.

3.5.14 Silt Density Index

Silt density index (SDI) test was based on the American Society For Testing Materials (ASTM D 4189-95) standard. Time it takes at 207 kPa to run a 500 mL sample of water through a 0.45 micron, 47 mm filter were recorded and compared with the time it takes to run another 500 mL sample after 15 minutes. The equation used for SDI calculation is:

$$SDI = 100(1 - (T_{initial} / T_{final})) / T \qquad \text{Equation 17}$$

$T_{initial}$ = Duration of time for initial 500 mL sample

T_{final} = Duration of time for final 500 mL sample

T = Time from start of $T_{initial}$ and start of T_{final}

Based on above equation, the maximum theoretical SDI attainable at 15 min is 6.7 as shown below:

SDI = 100 × (1-($T_{initial}$ / infinity))/ 15 = 100 × (1-0) / 15 6.7

So the method cannot be applied to influent water sample. The instrument used for this test was DigiSDI™ provided by TAKA™ Inc.

3.5.15 Scanning Electron Microscopy

Scanning Electron Microscopy (SEM) was based on "Fundamentals of Biological Scanning Electron Microscopy" (Chen, 1996). After the membrane surface was peeled off from the membrane tube, the interested parts were cut and fixed by 2.5% glutaraldehyde in Millonig's buffer, pH 7.3 for overnight at 4 °C. The specimens were postfixed in 1% OsO_4 in the same buffer for 2 hours in a refrigerator. They were washed briefly in distilled water, dehydrated in a graded series of ethanol solutions and absolute ethanol. Samples were then critical point dried, mounted on stubs and coated with gold in a sputter coater (Edwards S150B; Wdwards, West Sussex, England). Specimens were observed using a Hitachi S-2500 (Hitachi, Tokyo, Japan) scanning electron microscope.

4. Results and discussion

4.1 Permeate Water Quality

In this project, three types of water were used as feed water for the membrane tank: utility water, final effluent from GBWWTP that has been sterilized by UV and final effluent from the tank #6 of GBWWTP without the UV sterilization. The former one was named as FE1 and the latter one was named as FE2. FE1 was the combined water from the 10 tanks of GBWWTP. The impact of different feed waters on the permeate water quality by ZeeWeed® 500D membrane treatment and on the membrane performance was investigated.

The permeate water qualities achieved by ZeeWeed® 500D for FE1 and FE2 are summarized in Table 4.1 and 4.2, respectively. The water qualities for both feed waters are also included in Table 4.1 and Table 4.2, respectively. The feed and permeate water quality are displayed as the mean ± standard deviation, maximum, minimum values and percentage removal. Percentage removal was not shown for those parameters, where no decrease has been observed or elimination percentage could not be calculated. The raw water quality data of this project are presented in Appendix E.

By comparison of the water quality of these two feed waters shown in Table 4.1 and 4.2, there was no significant difference observed between them except the microorganisms. FE1 TSS was

4.3 ± 2.7 mg/L, while the FE2 TSS was 4.7 ± 2.6 mg/L; the FE1 TKN was 8.9 ± 4.7 mg/L, while the FE2 TKN was 2.5 ± 3.2 mg/L. In terms of coliform, the FE1 had much lower TC and FC value than FE2. The FE1 total coliform was 2358 ± 2334 CFU/100 mL; the FE2 total coliform was 210818 ± 256836 CFU/100 mL; the FE1 fecal coliform was 57 ± 47 CFU/100 mL; the FE2 fecal coliform was 17618 ± 30864 CFU/100 mL.

Permeates from the different feed waters have more common characteristics. TSS was less than the detection limit 0.6 mg/L; Turbidity was less than 0.2 NTU; BOD_5 was less than the detection limit 2 mg/L; coliform concentration was very low. There is no significant difference in permeate water quality for both feed waters. So the impact of different feed waters on the permeate water quality can be neglected. As mentioned above, the two feed waters have significant difference only in microorganisms that are usually too large to penetrate the ultrafiltration membrane. And the other properties for both feed waters are very similar. Therefore, the permeate for both feed waters should also be very similar, which is consistent with experimental results (permeate water quality shown in Table 4.1 and 4.2).

Table 4.1. The summary of FE1 (influent after UV) and permeate water qualities.

		Influent				Permeate			
Analytic parameters	Elimination percentage (%)	Mean	Standard deviation	Min	Max	Mean	Standard deviation	Min	Max
PH	-	7.5*	0.2	7.0	7.7	7.7	0.2	7.4	8.0
Temperature °C	-	13.5*		11.7	17.2				
Conductivity, uS/cm	-	995	80.2	838	1120	1001	116.1	827	1322
TSS, mg/L	H	4.3	2.7	1.8	16	<0.6			2
Turbidity, NTU	H	3.9	1.5	2.2	6.9	<0.2			
COD, mg O_2/L	35	36.8	5.4	24	52	24.0	3.6	16	35
BOD, mgO_2/L	H	3.7		2	13	<2			
TOC, mg/L	-	-				9	0.4	8	10
NH_4^+-N, mg/L	-	6.6	4.0	0.4	16.5	6.8	4.3	0.3	17.5
TOXN, mg/L	-	9.9	2.5	4.9	15.8	10.0	2.5	5.3	16.4
TKN, mg/L	8	8.9	4.7	1.9	20.5	8.4	4.9	1.4	20.7
TP, mg/L	34	0.7	0.4	0.3	2.0	0.5	0.3	0.1	1.2
TA, mg/L						176	21	125	223
TH, mg/L						228	31	162	281
Ca^{2+}, mg/L						49.9	7.6	21.1	84.5
Mg^{2+}, mg/L						42.3	6.0	13.2	54.0
Cl^-, mg/L						66.0	30.5	34.0	154.0
SO_4^{2-}, mg/L						123.6	16.2	92	150
HCO_3						183.1	19.4	137	211
SiO_2						7.9	0.4	7.1	8.6
TC colony/100mL		2358	2334	235	11000			<1	270
FC, colony/100mL		57	47	8	220			<1	2

Note: The result would be discussed in detail.

Table 4.2. The summary of FE2 (influent before UV) and permeate water qualities.

		Influent				Permeate			
Analytic parameters	Elimination percentage (%)	Mean	Standard deviation	Min	Max	Mean	Standard deviation	Min	Max
PH		7.2	0.3	6.8	7.6	7.6	0.5	6.4	8.2
Temperature °C									
Conductivity, uS/cm		1022	354.4	852	1204	1019	353.0	858	1204
TSS, mg/L	H	4.7	2.6	0.8	13	<0.6		<0.6	1.2
Turbidity, NTU	H	2.6	0.5	1.9	3.3	<0.2			
COD, mg O_2/L	28	34.4	8.0	21	61	23.5	5.5	9	37
BOD, mgO_2/L	H	-	-	<2	14	<2		<2	3.6
TOC, mg/L		11.4	1.2	9	13	10.6	1.3	9	13
NH_4^+-N, mg/L		0.5	1.8	0.004	10.3	0.5	1.9	0.004	10.8
TOXN, mg/L		5.4	3.2	0.15	15.6	5.6	2.7	0.2	12.1
TKN, mg/L	32	2.5	3.2	1.2	17.8	1.9	2.9	0.8	15.8
TP, mg/L	36	0.8	1.0	0	4.6	0.6	1.0	0.07	4.2
TA, mg/L						189.8	24.7	122	259
TH, mg/L						311.8	55.5	222	462
Ca^{2+}, mg/L						65.7	13.6	51	97.3
Mg^{2+}, mg/L						27.2	6.4	18	40
Cl^-, mg/L						51.4	2.8	44.7	55.6
SO_4^{2-}, mg/L						212.5	44.3	170	300
HCO_3						189.8	24.7	122	259
SiO_2		9.1	0.6	8.2	10.3	9.0	0.7	8.2	10.3
TC colony/100mL		210818	256836	15000	1500000			<1	330
FC, colony/100mL		17618	30864	1100	170000			<1	11

Note: The result would be discussed in detail.

4.1.1 Solids Removal

As expected, the permeate TSS should be less than the detection limit and independent on the variation of solids loading to the membrane tank, because TSS is measured by the weight increased on glass-fiber filter, which has bigger pore size than the ZeeWeed 500D® membrane pore size (0.04 μm). Therefore all the contaminants, which cannot be retained by ZeeWeed 500D® membrane, should not be retained by the glass-fiber filter.

Figure 4.1 and 4.2 show the variation of TSS in the influent and permeate with time. As shown in these two figures, the TSS for most of the influent samples is around 4 mg/L, however, with a large variation from 0.8 to 16 mg/L. More than 80% of permeate samples have a very low TSS, less than 0.6 mg/L, exhibiting a high solids removal efficiency. The permeate TSS results indicate that suspended solids in feed waters were mostly rejected by this ultrafiltration membrane during this experimental period.

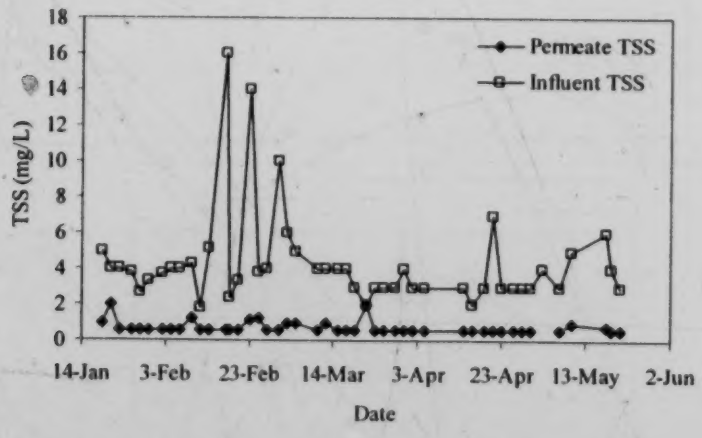

Figure 4.1. Variation of TSS when FE1 used as the feed water.

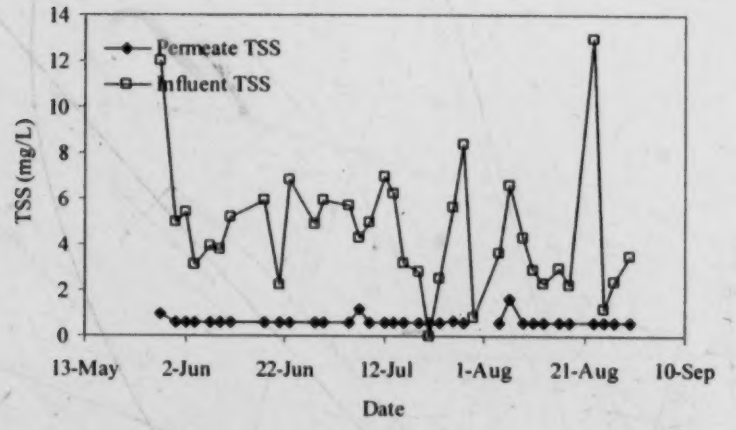

Figure 4.2. Variation of TSS when FE2 used as the feed water.

4.1.2 COD Removal

Compared to the feed water, a decrease of organic matter in permeate was consistently observed during the long run period for both feed waters. Average 35% and 28% COD removal efficiency were achieved by the ZeeWeed 500D® membrane for FE1 and FE2, respectively (Table 4.1 and Table 4.2). As shown in Figure 4.3 and 4.4, the influent COD values in the long-term experiment ranged from 24 to 52 mg/L for FE1 with an average 35 mg/L and from 21 to 61 mg/L for FE2 with an average 34 mg/L. The permeate COD was not significantly reduced ranging from 15 to 35 mg/L with an average 24 mg/L for FE1 as feed water and from 15 to 30 mg/L with an average 23.5 mg/L for FE2 as feed water. Figure 4.4 also shows that the permeate COD varied with the influent COD. When the influent COD was higher, the permeate COD was higher.

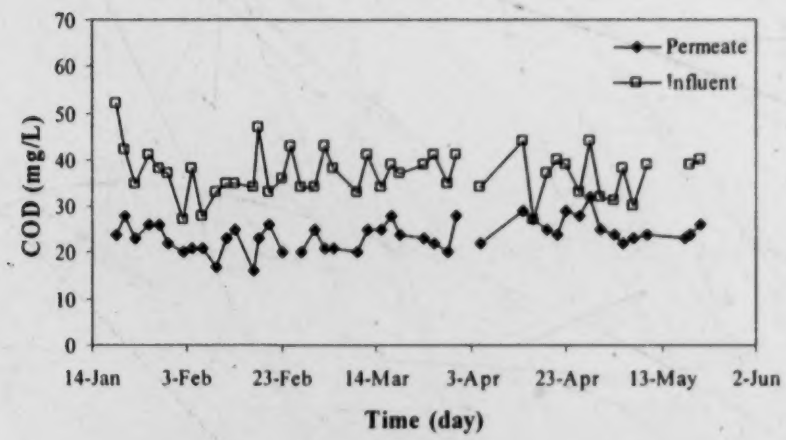

Figure 4.3. Variation of COD when FE1 used as the feed water.

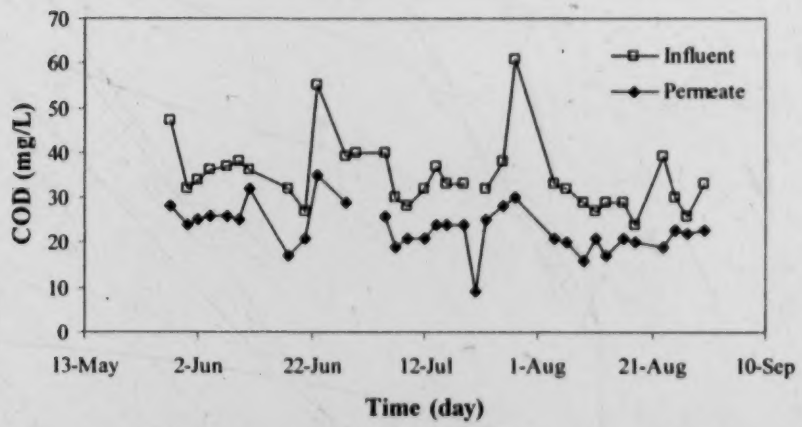

Figure 4.4. Variation of COD when FE2 used as the feed water.

4.1.3 Nitrogen Removal

NH_4^+-N, TOXN and TKN were analyzed in this project. Table 4.1 and Table 4.2 indicate that there were no significant removals for NH_4^+-N and TOXN. In case of TKN, only 8% of TKN was removed when FE1 was used as feed water and 32% was removed when FE2 was used as feed water, but the removal amount were almost the same ~0.7 mg/L, for both feed waters despite

different removal efficiency for two feed waters.

The pore size of ZeeWeed® 500D membrane was 0.04 μm, much larger that the size of the molecular size of some ions, therefore this membrane cannot be used to reject ion contaminants. So the fact that there is no considerable removal for NH_4^+-N and TOXN is within the expectation. However, TKN includes NH_4^+-N and organic nitrogen, which probably associates with macromolecules. Some of those macromolecules whose molecular sizes are bigger than the pore size of the membrane can be retained by the membrane. The other macromolecules smaller than the membrane pore size will pass through the membrane. So certain removal for TKN was expected.

The reason for the different removal percentage between the FE1 used as feed water and FE2 used as feed water, is attributed to the different nitrogen component of TKN. From Jan 19, 2004 to May 28, 2004, the final effluent after UV (FE1) was used as the feed water. During that period, Edmonton area was in the wintertime. The performance of the whole GBWWTP during wintertime was different from summertime during the May 28, 2004 to October 3, 2004 while the final effluent before UV (FE2) was used as the feed water. Also, the FE1 was the composite of 10 BNR tank in the GBWWTP, while the FE2 was only the effluent of tank 6 in the GBWWTP. Both the different season and different feed water source could cause the different TKN removals.

4.1.4 Phosphorus Removal

The total phosphorus removal was around 35% in the nine months of operation regardless of the feed water and operating conditions. But from the Table 4.1 and 4.2, the amount eliminated was only around 0.2 mg/L. So, the phosphorus mass removal was not very significant in this project even though a relatively good TP removal efficiency was achieved; only the part of phosphorus associated with suspended particulates or macromolecules was rejected.

4.1.5 Metal Removal

The metal concentrations in the influent, permeate and overflow were measured three times (on May 5th, May 7th and May 10th), when the FE1 was used as the feed water. Selected metal analysis results are listed in Table 4.3 with the complete results in Appendix E.

There was a significant elimination of the following metals: aluminum, cadmium, chromium, copper and lead. It was also found that these metals were concentrated in the overflow. All of these information indicate that aluminum, cadmium, chromium, copper and lead are rejected by the ZeeWeed® 500D membrane. The rejection efficiencies were 29%, 17%, 12%, 56%, and 13%, respectively. Occasionally, some other metals were also reduced by a considerable amount. This is consistent with the previous work (Alonso et al. 2001), except that the different metals were found to be retained.

Table 4.3. The reduction of metals.

Iron name	Average reduction[A] (%)	Average concentrate[B] (%)
Aluminum	29	161
Cadmium	17	125
Chromium	12	114
Copper	56	156
Lead	13	113
Manganese	5	109
Titanium	5	118
Zinc	5	105

[A] (Influent metal's concentration – permeate metal's concentration) / Influent metal's concentration × 100 %

[B] Influent metal's concentration / Overflow metal's concentration × 100 %

4.1.6 Microorganism Reduction

Coliform samples were collected three times per week. The TC and FC data in the membrane tank are summarized in Table 4.4. All the total coliform and fecal coliform data are presented in the Figures 4.5, 4.6, 4.7 and 4.8.

Table 4.4. Summary of coliform in the membrane tank.

	FE1 as feed water		FE2 as feed water	
	Average	Standard deviation	Average	Standard deviation
TC (cfu/100 mL)	1.15×10^5	5.72×10^5	8.17×10^5	1.39×10^6
FC (cfu/100 mL)	290	555	4.18×10^4	8.43×10^4

Figures 4.5, 4.6, 4.7, and 4.8 show that most TC and FC values for permeate were less than 1 CFU/100 mL except some of the data were randomly high. There was no clue that these high TC and FC data have certain relation with the membrane operating condition. These figures also show that the performance of ZeeWeed® 500D membrane for the high microorganism reduction

was independent of the feed coliform concentration, and microorganisms were almost 100% rejected by the membrane no matter how high the feed water count of microorganisms was. This further proved that the size of microorganisms were larger than the pore size of membrane, as discussed in Part 4.1.

The random high coliform data for permeate could not support the opinion that membrane integrity was an issue because three times of bubble test during this project did not show membrane leakage. The expected explanation for the high coliform data is sample contamination. There are two suspected contamination sources: the permeate tank and air. As shown in Figure 3.2, the permeate tank was used as the backpulse tank in this experiment design and permeate kept overflowing from the backpulse tank, so backpulse tank could not be sterilized by chemicals during the operation. With time, some microorganisms could grow on the backpulse tank wall and some of them could go to the membrane system with the back pulse water and contaminate the permeate water randomly. Some microorganisms were indeed found growing on the backpulse tank wall during the project (Figure 4.9). On May 17th, some NaOCl was dosed to the backpulse tank and left to stay overnight. Figure 4.9 is the picture of the backpulse tank wall before chemical cleaning and Figure 4.10 is the picture of the backpulse tank after the chemical cleaning. It was apparent from these two pictures that the backpulse tank was cleaned by the chemical treatment. After the chemical cleaning, less TC and FC were found in the permeate samples, but random TC data were shown again several days later.

Another possible contamination source is the air. The aerosol coliforms inside and outside of the membrane tent were tested on August 24, 2004 (Table 4.5). These data shows the average aerosol coliform inside of the membrane tent was 438 CFU/m^3, while the average value of the outside of the membrane tent was 43 CFU/m^3 after 24 hours incubation. The high aerosol coliform concentration in the membrane tank increases the chance of contamination during the sampling process. The aerosol coliform concentration inside the membrane tent could be affected by temperature and ventilation condition, so it was very difficult to find the relationship between the coliform in permeate water samples and aerosol coliform concentration in the membrane tent. For the safety of the operator, it was suggested to keep the membrane tent in good ventilation condition.

Table 4.5. Comparison of aerosol coliform inside and outside of the membrane tank.

Coliform (CFU/m^3)	After 24 h incubation	After 72 h incubation
Inside of membrane tank	438	535
Outside of membrane tank	43	152

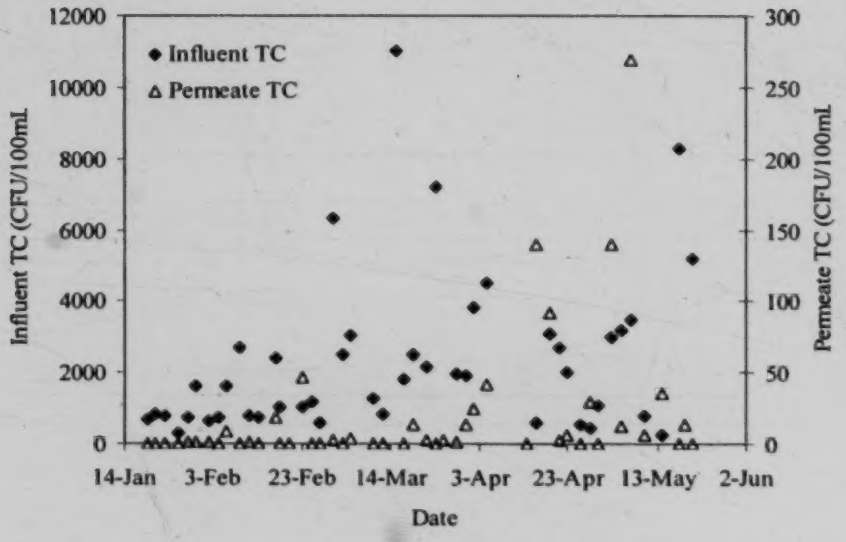

Figure 4.5. Permeate and influent TC for FE1 as the feed water.

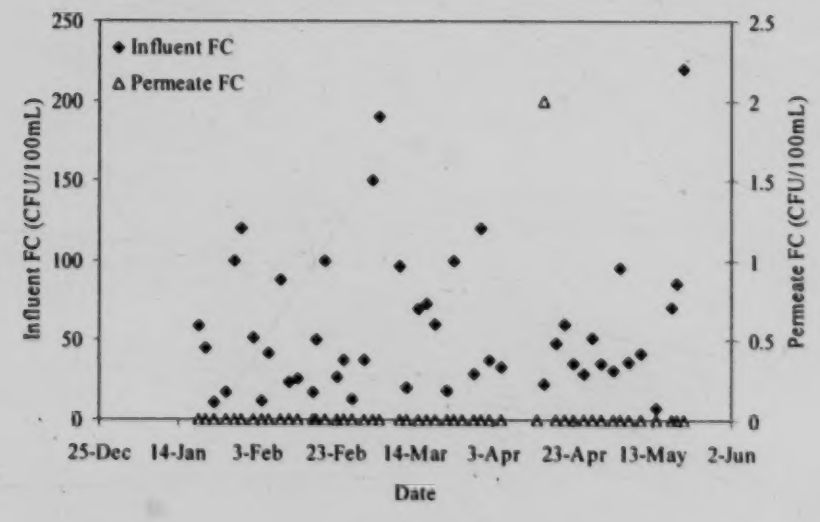

Figure 4.6. Permeate and influent FC for FE1 as the feed water.

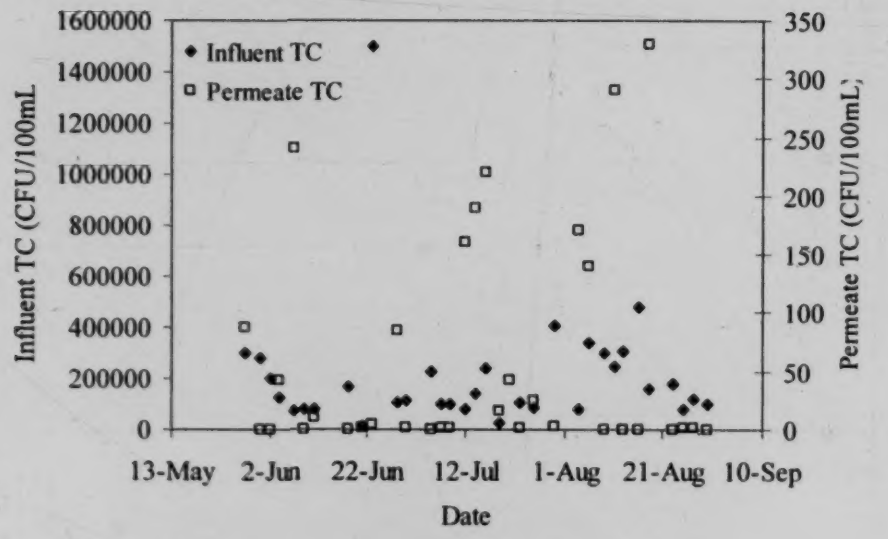

Figure 4.7. Permeate and influent TC for FE2 as the feed water.

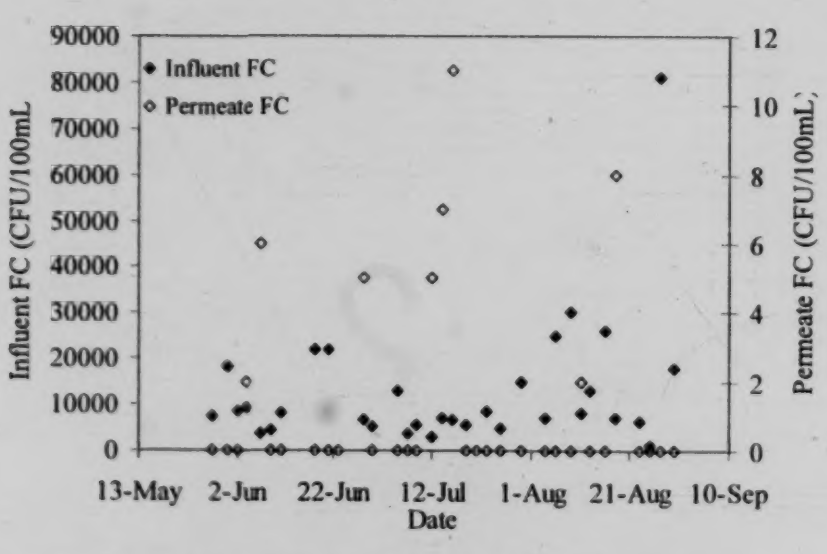

Figure 4.8. Permeate and influent FC for FE2 as the feed water.

Figure 4.9. Backpulse tank before chemical cleaning.

Figure 4.10. Backpulse tank after chemical cleaning.

4.1.7 Coliphage Test

The influent, permeate and overflow water samples were tested for coliphage. Due to the different coliphage concentration, 100 mL water sample was collected for the influent and permeate, and only 2 mL sample was collected for the overflow. Totally, there were 20 permeate samples and 32 overflow samples tested. Two influent samples were tested for each test of corresponding permeate and influent sample.

The 20 permeate samples were collected from the same operating parameter set points (15 minutes operating time, 30 seconds backpulse, 23.7 $Lm^{-2}h^{-1}$ flux). Four samples were collected in one operating cycle in July 19, 5 samples in August 18 and 5 samples in August 23. The results were presented in Table 4.6. The highest coliphage concentration of the 20 samples was 5 CFU/100 mL. The coliphage concentration in the influent varied from 384 CFU/100 mL to 1200 CFU/100 mL. ZeeWeed® 500D membrane shows a very high log removal for coliphage regardless of the feed concentration.

Table 4.6. Coliphage in permeate and influent water samples

Time after BP, July 19	48 sec	340 sec	640 sec	840 sec		Influent
Coliphage (CFU/100 mL)	0	5	1	0		
Time after BP, August 18	100 sec	300 sec	500 sec	700 sec	800 sec	
Coliphage (CFU/100 mL)	2	1	0	1	1	384
Time after BP, August 23	100 sec	200 sec	500 sec	700 sec	800 sec	
Coliphage (CFU/100 mL)	0	2	0	0	0	1200

Figure 4.11 shows the coliphage concentration in the membrane tank against one operating cycle time. Time zero is the end of last backpulse and the beginning of a new cycle. The x-axis is the time after last backpulse. The coliphage concentration in the membrane tank reached a peak point just seconds after the last backpulse finished.

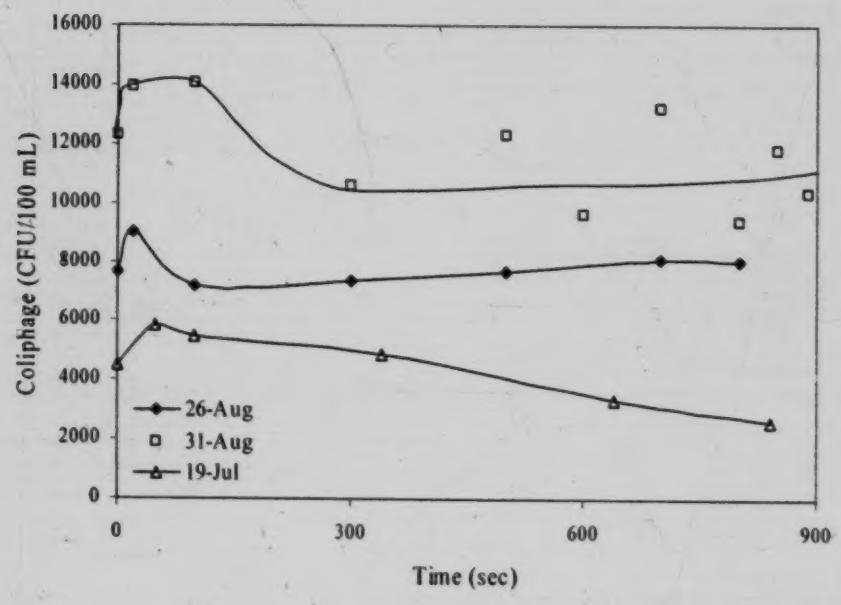

Figure 4.11. Coliphage concentration against time in one operating cycle.

During the production time, coliphages were almost completely rejected by the membrane. Some of them accumulated on the membrane surface and some of them concentrated in the membrane tank. In the immediate following backpulse, the coliphages accumulated on the membrane surface

would be flushed off by the backpulse flow and the coliphage concentration in the membrane tank was then increased. Then the coliphage concentration in the membrane tank reached the peak point. However, the permeate water was reversed back to the membrane tank at the speed of 8 L/s during the backpulse; also the influent was still entering the membrane tank during this time, so the coliphage concentration in the membrane tank decreased quickly after the peak point. In the immediate production period, coliphage concentration in the membrane tank should slightly increase gradually since coliphage in membrane tank was concentrated somewhat by membrane. But the results shown in Figure 4.11 do not exhibit the gradually slight increase for all three operating cycles.

4.1.8 Silt Density Index

In this project, certain impact of the membrane fouling conditions and membrane operating parameters on the permeate water quality was expected. But the results of particle counter, TSS and coliphage test did not show this type of pattern, due to the limitation of these methods. Finally, silt density index (SDI) test was found to be able to detect the difference of permeate water quality caused by these factors.

In the Figure 4.12, the samples were collected for two different fluxes: 23.7 $Lm^{-2}h^{-1}$ and 19.0 $Lm^{-2}h^{-1}$, but in the same day, which ensured a similar membrane irreversible fouling condition for all collected samples. Due to large amount of water sample (20 L) required for SDI test, only two samples were collected in each operating cycle: one was collected at the beginning of operating

cycle and the other one was collected at the end of the operating cycle just before the next backpulse. These two samples were collected in the same operating cycle, but at different membrane reversible fouling conditions, as can be imagined by referring to the later discussion. This figure shows that the two SDI data for flux at 23.7 $Lm^{-2}h^{-1}$ are higher than the values at corresponding time for flux at 19.0 $Lm^{-2}h^{-1}$. This can be explained by the different TMP and suction pressure. When higher flux was produced, higher TMP and suction pressure were required, therefore, leading to higher SDI value. Figure 4.12 and Figure 4.13 clearly illustrate this finding.

Figure 4.12 also shows that the SDI in one operating cycle decreases with time. This can be explained by the reversible fouling including the cake layer formed on the membrane surface and the solids sucked in the membrane pores during the operating cycle. The cake layer formed on the membrane surface acts as another filter and solids sucked in the membrane pores reduce the membrane pore size. Both of them help the ZeeWeed 500D to produce higher permeate water quality. Ahn and Song (2000) proposed a similar assumption, but they did not have experimental data to prove it.

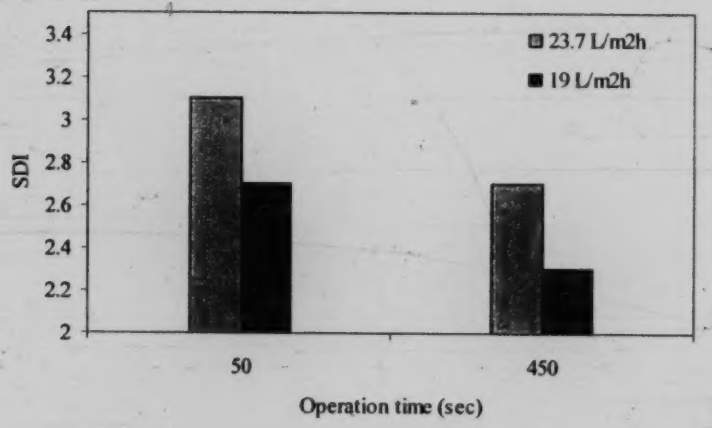

Figure 4.12. The effect of flux and reversible fouling on SDI.

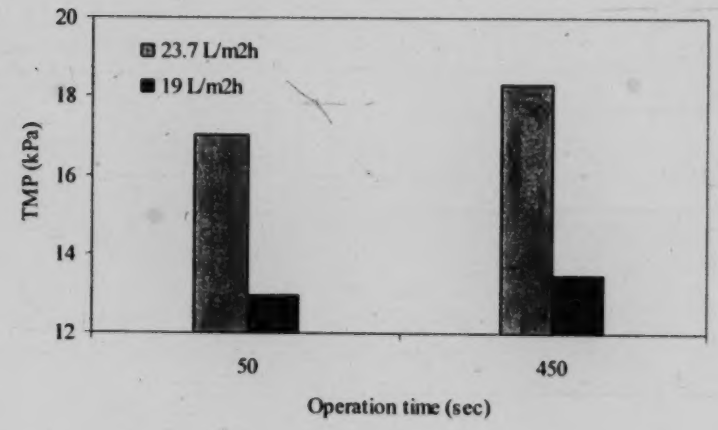

Figure 4.13. TMP when the SDI samples collected.

The cake layer and solids blocked in membrane pores in Figure 4.14 are gradually formed in one operating cycle. Most part of them (reversible fouling) will be blown off by the following backpulse, but small part of them won't be removed by backpulse, which is the irreversible fouling. When more and more irreversible fouling accumulates on the membrane surface, higher TMP will be required to keep the same flux, which will increase the absolute suction pressure and

make more solids pass through the membrane. So the impact of the cake layer on the permeate water quality is double-faced.

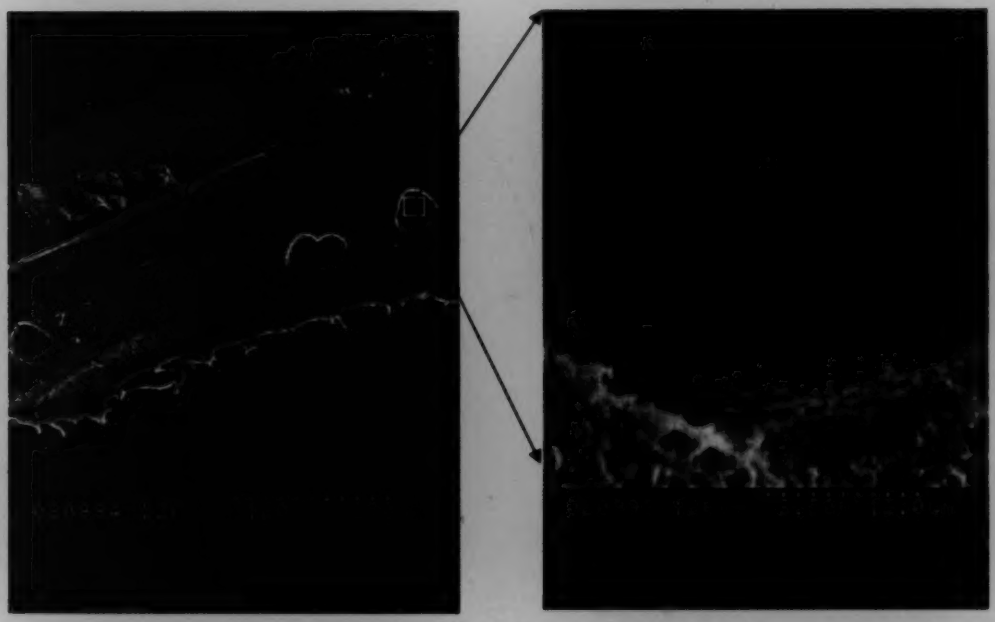

Figure 4.14. SEM pictures of cake layer on the membrane surface and solids drawn into the membrane pores.

(Note: The membrane samples for scanning electron microscope (SEM) pictures were taken several days after chemical cleaning and just at the end of one operating cycle. So both the reversible fouling and irreversible fouling were very apparent.)

Figure 4.15 illustrates the complex impact of membrane fouling on the permeate SDI. The results span from the first day after maintenance cleaning till the end of the operation. Each day two samples were collected from one operating cycle. One sample was collected at the beginning of operating cycle and the other one was collected at the end of the same operating cycle. Also, the TMP of membrane system was recorded when samples were collected (Figure 4.16).

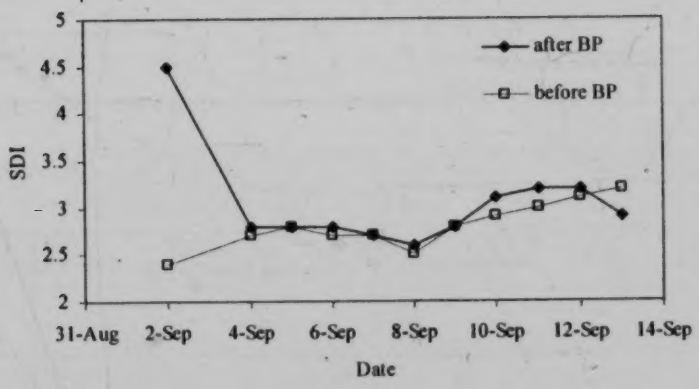

Figure 4.15. SDI data versus long-term operation.

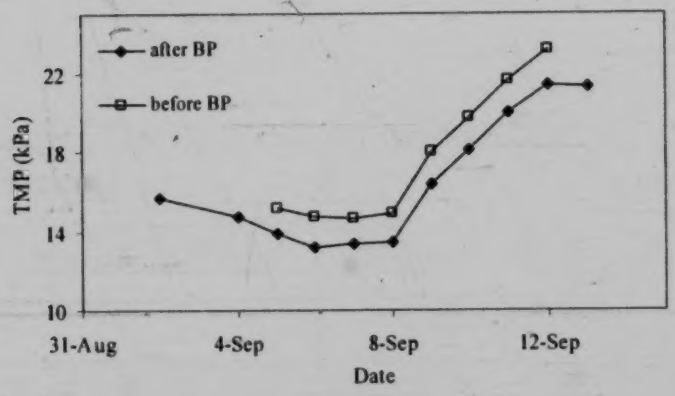

Figure 4.16. TMP versus long-term operation.

At beginning of this test, it was assumed that the membrane surface was very clean and there was no irreversible fouling. The irreversible fouling gradually accumulated on the membrane surface with time. The TMP at the beginning of each operating cycle was used to indicate the extent of the irreversible fouling. The TMP at the end of the each operating cycle was used to indicate the extent of the reversible fouling.

The SDI curve for "after backpulse" began with 4.5, decreased with time in the first six days and then increased after the 6th day till the end of this test. The high SDI value at the beginning of this test was attributed to the fact that the membrane surface was cleaned and there was almost no irreversible fouling after the chemical cleaning. However, some irreversible fouling began to accumulate on the membrane surface after the first day, but this did not increase the TMP. So the SDI began to decrease due to the help from the irreversible fouling. After 6 days, the TMP for "after backpulse" began to increase, which was caused by increasing irreversible fouling. Since then, the SDI curve for "after backpulse" began to increase.

The SDI curve for "before backpulse" also reached the lowest point at the 6th day and began to increase after that day. Figure 4.16 shows the TMP curve for "before backpulse" kept horizontal within the first six days and increased after six days. So the increase of the SDI probably was caused by the higher TMP.

4.2 Operating Factors

To optimize the operating parameter set points, five factors were tested and monitored in this project, including the frequency of backpulse (operating time), duration of backpulse, net permeate flux, aeration amount and water recovery. Normally, four factors were kept fixed and only one factor was adjusted to figure out the impact of the adjusted factor on membrane performance. However, sometimes the impacts of these factors were very complicated and entangled, which are demonstrated in the following discussion. In the following part, the impact of the five factors is separately discussed. Finally, a safe operating range is developed to predict the membrane performance under different operating set points.

4.2.1 Flux

As the flux increased, the accumulation of deposit progressed rapidly (Ahn and Song, 2000) . But how the flux impact the fouling and membrane performance is different for different membranes. Figure 4.17, 4.18 and 4.19 show the impact of flux on HTMP for different operating times (15 min, 25 min and 10 min), respectively. All the seven tests were run under the same aeration amount (237 m^3h^{-1}), the same water recovery (70%) and the same backpulse duration (30 seconds). As discussed in part 3, the HTMP can be used as a signal of fouling condition in this membrane system.

Figure 4.17 shows that the ZeeWeed® 500D membrane system can run 106 hours or even longer without obvious fouling with the flux set point at 23.7 $Lm^{-2}h^{-1}$ for 15 minutes operating time.

However, at a high flux of 28.4 $Lm^{-2}h^{-1}$ for the same operating time, fouling was clearly observed even from the beginning of the test and the membrane system was shutdown after 24.5 hours due to the severe fouling (HTMP increased up).

As shown in Figure 4.18, for 10 minutes operating time the increase of flux from 23.7 $Lm^{-2}h^{-1}$ to 28.4 $Lm^{-2}h^{-1}$ did not cause worse membrane performance. Membrane system could keep running for a long time, except that the average HTMP was 5 kPa higher under 28.4 $Lm^{-2}h^{-1}$ flux than that under 23.7 $Lm^{-2}h^{-1}$. The higher HTMP, the more fouling happened in one operating cycle. However, the shorter operating time 10 minutes (more frequent backpulse) can effectively clean the fouling and keep the system running for 95 hours without shutdown.

Figure 4.19 is the HTMP versus time for 25 minutes operating time. It shows that the membrane system could not run longer than 20 hours with flux at the 23.7 $Lm^{-2}h^{-1}$ and not longer than 35 hours for 19.0 $Lm^{-2}h^{-1}$ flux. However, when the flux was reduced to 14.3 $Lm^{-2}h^{-1}$, the system could run 160 hours without shutdown.

Once the system shut down due to the severe fouling, chemical cleaning was required. However, the objective of the membrane operation is highest operating flux with lower frequency of chemical cleaning. So a practical operating flux with an acceptable chemical cleaning frequency was expected. When the chemical cleaning was operated once in a period of two weeks, the practical operating flux for 15 minutes operating time was below 23.7 $Lm^{-2}h^{-1}$; the practical

operating flux for 10 minutes operating time was below 28.4 $Lm^{-2}h^{-1}$; the practical operating flux for 25 minutes operating time was below 14.3 $Lm^{-2}h^{-1}$. It were concluded that more frequent backpulse supported higher flux. However, more backpulse will consume more permeate water. In another word, more backpulse will decrease the flux in the long term. So the balance between the frequency and flux should be considered.

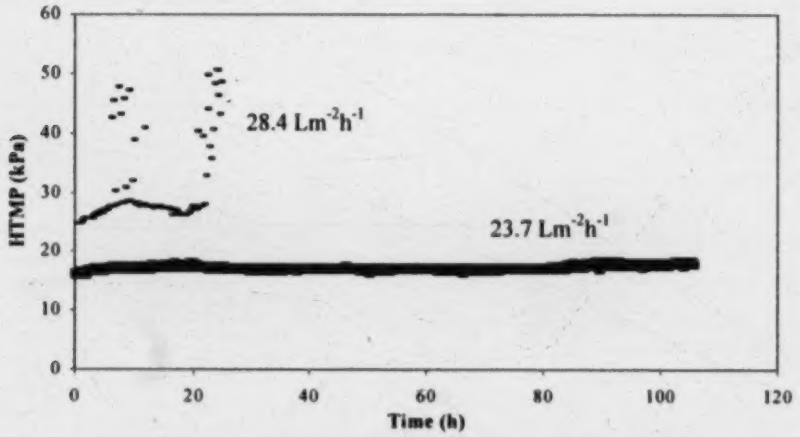

Figure 4.17. HTMP versus time under 15 minutes operating time.

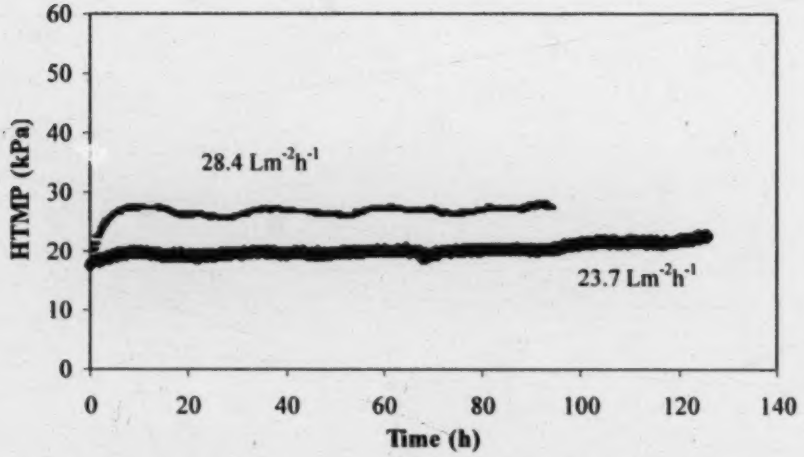

Figure 4.18. HTMP versus time under 10 minutes operating time.

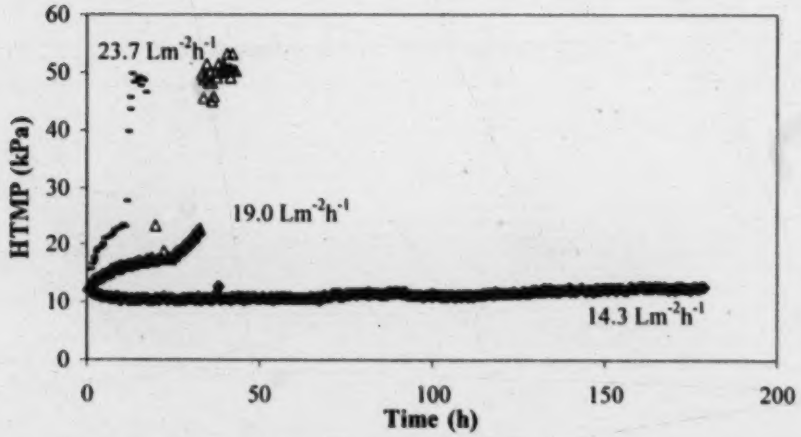

Figure 4.19. HTMP versus time under 25 minutes operating time.

4.2.2 Operating time

Figure 4.20 shows the impact of the operating time on membrane performance with the same flux, backpulse duration, aeration amount and water recovery set points. Shorter operating time means higher frequency of backpulse. So 10 minutes operating time has more backpulse compared with

25 minutes operating time in the same period of time. This figure shows that HTMP of 25 minutes operating time increased quickly due to the incomplete fouling recovery and reached the system's highest tolerable TMP 50 kPa after 16.25 hours, and then the system shut down automatically. For 10 minutes and 15 minutes operating time, the system had a very good performance in a period of more than 100 hours. Compared to 10 minutes operating time, 15 minutes operating time had a lower operating pressure . This is because 10 minutes operating had more frequent backpulse and more permeate water was consumed by the backpulse. In order to compensate the permeate water loss due to backpulse and reach the average flux of 23.7 $Lm^{-2}h^{-1}$ during one operating cycle, the system had to automatically increase the actual flux during the permeate water production period. Therefore, the actual flux during production period for 10 minutes operating time was higher than that for 15 minutes operating time, resulting in a higher TMP for 10 minutes operating time than 15 minutes operating time. So, compared with 15 minutes operating time, 10 minutes operating time required a higher flux and higher TMP, but both of them were acceptable.

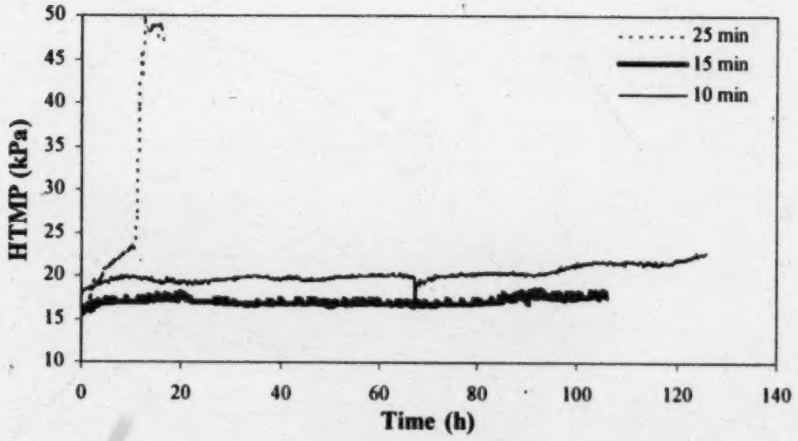

Figure 4.20. HTMP versus time under flux 23.7 L/m^2/h and BP 30 sec.

4.2.3 Duration of Backpulse

The high and stable specific flux in a long term means that the backpulse cleaning was effective. The low and decreased specific flux means the failure of the backpulse. To compare the effect of the different backpulses, the long-term specific fluxes after each backpulse were put together in Figure 4.21.

It was observed that the 30 seconds backpulse caused the highest specific flux and the specific flux kept stable in the long-term operation. The specific flux for the15 seconds backpulse and the 40 seconds backpulse were lower than 30 seconds backpulse and tended to decrease with time. So 15 seconds backpulse turned out to be too short to remove the cake layer on the membrane surface and should not be applied in this membrane operation. The 40 seconds backpulse was supposed to have higher cleaning effectiveness than 30 seconds backpulse. In contrast, lower

specific flux was observed. The reason for this is that the effectiveness of backpulse was affected by the thickness and denseness of the cake layer. In these three tests, the flux set point was the same as 23.7 $Lm^{-2}h^{-1}$, but the actual flux was not the same. Because the PLC system was automatically trying to compensate the water used for the different backpulses and caused different actual flux in three cases, even though the flux set point was the same. The longer backpulse duration, the higher the actual flux and the thicker the cake layer. At the same time, the higher actual flux requires higher TMP, which increase the density of cake layer. So the test with 40 seconds backpulse produced a thicker and denser cake, which influenced the effectiveness of backpulse. Figure 4.22 indicates the 40 seconds backpulse has the highest actual flux among the three cases. And Figure 4.23 shows that the highest TMP was required for the 40 seconds backpulse.

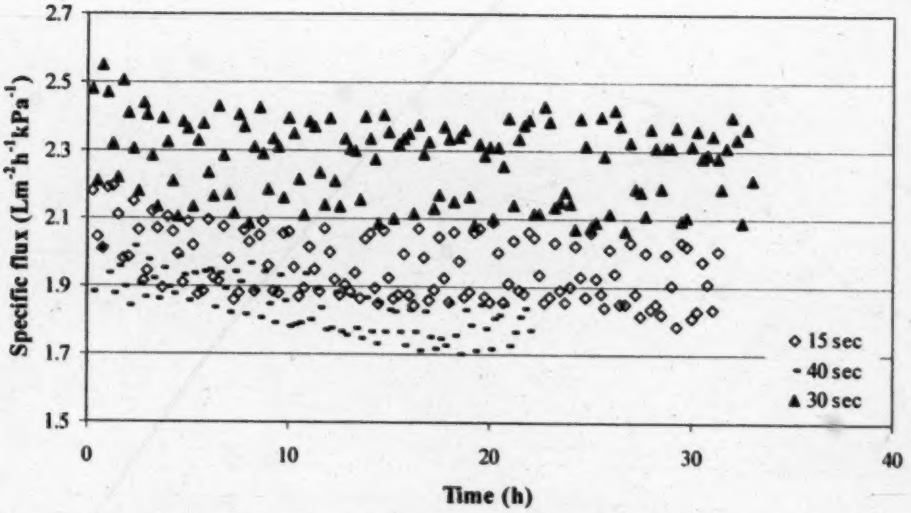

Figure 4.21. The impact of backpulse duration on specific flux.

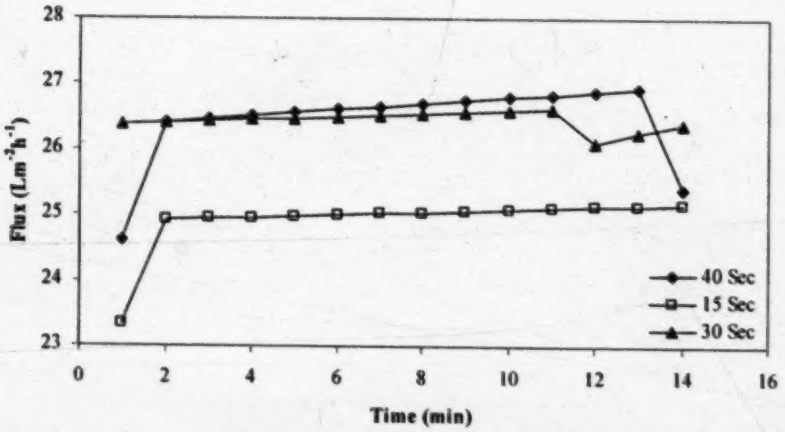

Figure 4.22. Actual flux versus time in one operating cycle with the same flux set point but different backpulse duration.

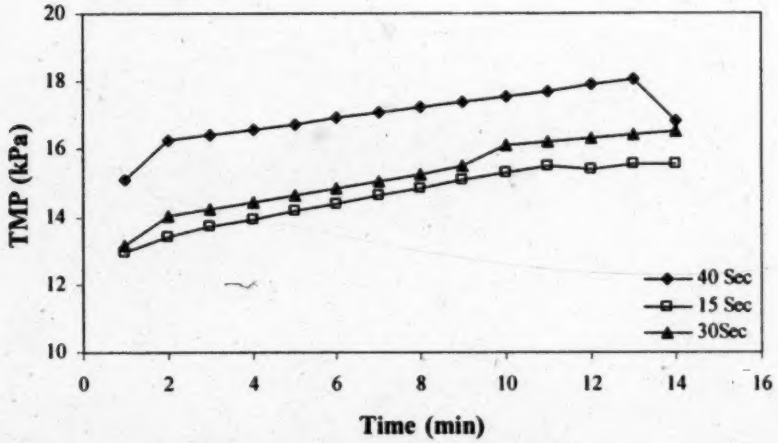

Figure 4.23. TMP versus one operating cycle with the same flux set point but different backpulse duration.

4.2.4 Aeration

In this experiment, aeration in the membrane tank was used to mix the membrane tank and blow the membrane surface to reduce the speed of cake layer building up. Two set points 203 m^3h^{-1} and 237 m^3h^{-1} were tested to check the impact of aeration. The results with different operating set points are presented in Figure 4.24 to Figure 4.26.

When the operating time was 15 minutes and 25 minutes, the runs with lower amount of aeration had a faster cake building up in each operating cycle, which was exhibited in a higher operating HTMP. When the operating time was 10 minutes, the impact of different aeration was not so significant.

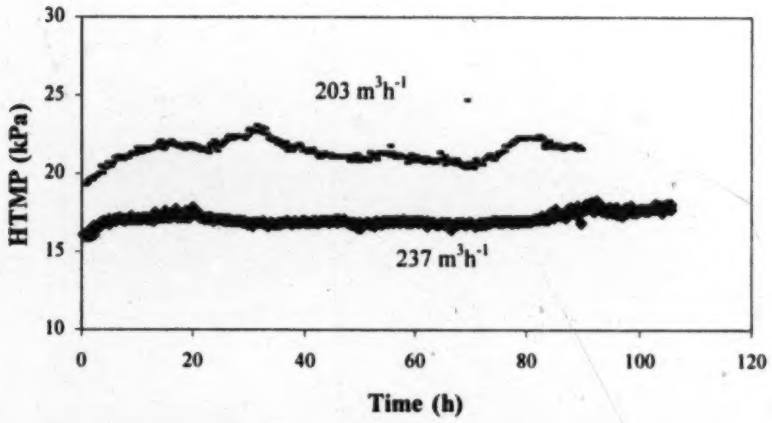

Figure 4.24. Impact of different aeration amounts with the same other operating set points (operating time15 min, flux 23.7 $Lm^{-2}h^{-1}$, backpulse 30 seconds).

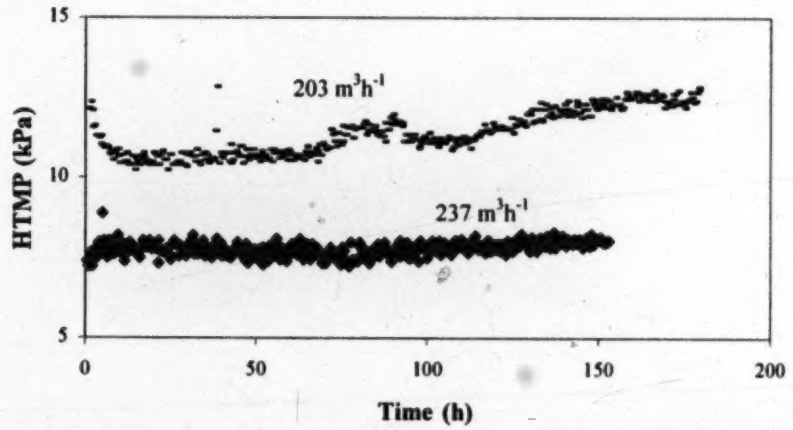

Figure 4.25. Impact of different aeration amounts with the same other operating set points (operating time 25 min, flux 14.3 $Lm^{-2}h^{-1}$, backpulse 30 seconds).

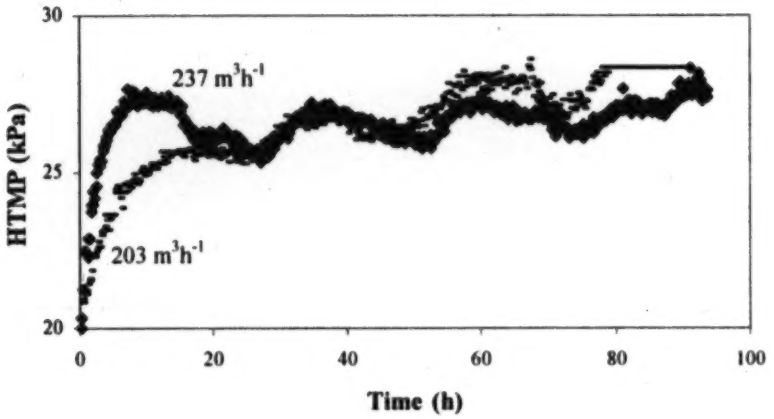

Figure 4.26. Impact of different aeration amounts with the same other operating set points (operation time 10 min, flux 28.4 $Lm^{-2}h^{-1}$, backpulse 30 seconds).

4.2.5 Water Recovery

Water recovery was used as another impact factor for the membrane performance. Different water recovery causes different TSS concentration in the membrane tank with the same influent TSS concentration. The different membrane tank TSS concentration will affect the speed of membrane fouling, indicated by different operating pressure. However, Figure 4.27 shows the HTMP curves with three different water recoveries were almost same. In this experiment, the membrane system was used as the tertiary treatment and influent TSS concentration was very low, so the impact from the water recovery was negligible in this experiment.

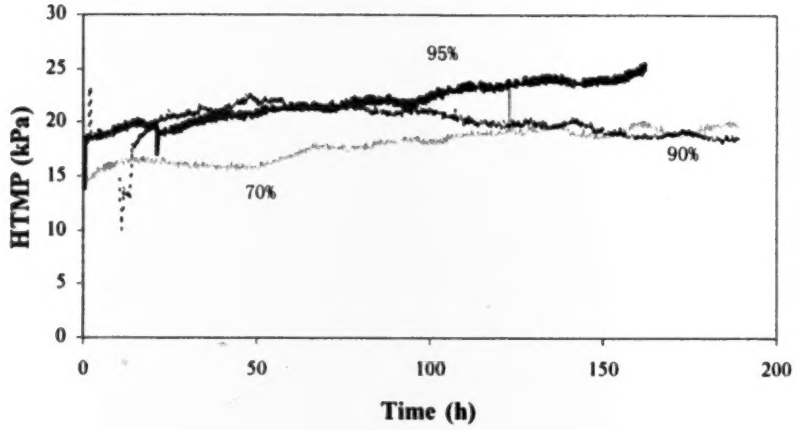

Figure 4.27. HTMP versus time with different water recovery.

4.2.6 Impact of Different Feed Water

FE1 was used as feed water for the membrane tank since January 19, 2004 to May 28, 2004. After that, FE2 was be used as feed water till October 3, 2004. The impact of different feed waters was tested based on the same operating parameter set points. In Figure 4.28 the operation conditions were set at 10 minutes operating time, 30 seconds backpulse, 23.7 $Lm^{-2}h^{-1}$ flux, 237 m^3h^{-1} aeration, and 70% water recovery. In Figure 4.29 the operating conditions were set at 15 minutes operating time, 30 seconds backpulse, 23.7 $Lm^{-2}h^{-1}$ flux, 237 m^3h^{-1} aeration and 70% water recovery. The HTMP curves for two feed waters as shown in these two figures were quite similar. So the impact of these two feed waters was negligible.

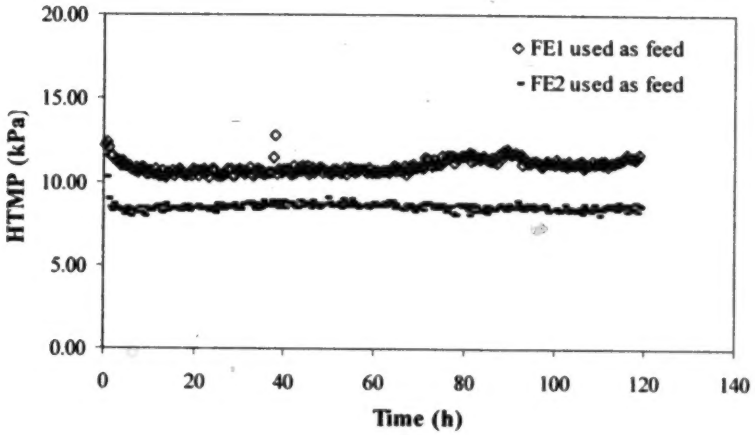

Figure 4.28. HTMP versus time with different feeds (operating time 10 minutes, backpulse 30 seconds, flux 23.7 $Lm^{-2}h^{-1}$, aeration 237 m^3h^{-1}, water recovery 70%).

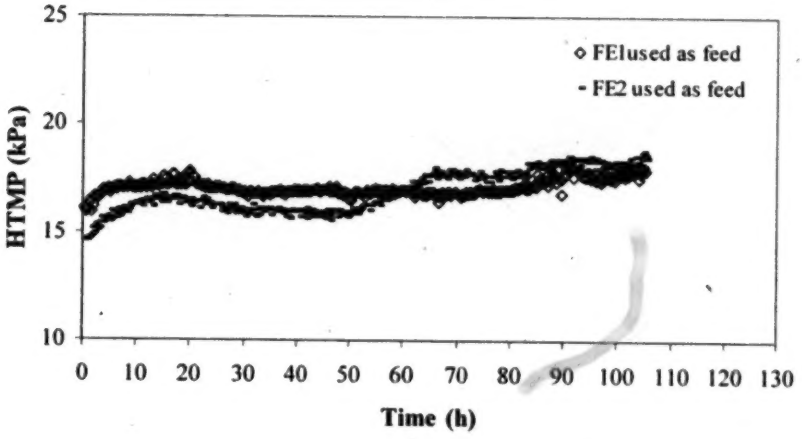

Figure 4.29. HTMP versus time with different feeds (operating time 15 minutes, backpulse 30 seconds, flux 23.7 $lm^{-2}h^{-1}$, aeration 237 m^3h^{-1}, water recovery 70%).

4.3 Membrane Operating Parameters Range

In the near future, ZeeWeed® 500D membrane will be scaled up and used as a tertiary treatment in GBWWTP. The ranges of operating parameters are required in advance for design. For the duration of backpulse, it has been shown that 30 seconds was the most effective time period. Water recovery and aeration amount was shown to slightly affect the membrane operating pressure, but did not impact the membrane long-term operating performance. So these three factors should be kept the same in the predication of other operating parameters.

All the data in the Figure 4.30 were taken from operating cycles with different flux value or different operating times, but with the same backpulse duration (30 seconds), aeration (237 m^3h^{-1}) and water recovery (70%). The y-axis is the average flux in 24 hours, which has been corrected to 20 °C. The x-axis is the average TMP in 24 hours. It was found that the flux values with the same operating time were proportional to the TMP. The slopes of these straight lines are very close.

The shaded area in Figure 4.30 is the practical operating range for flux and TMP, when ZeeWeed® 500D membrane is used for tertiary treatment. In this area the membrane flux range is between 16.5 $Lm^{-2}h^{-1}$ and 33.4 $Lm^{-2}h^{-1}$; TMP range is between 10.0 kPa and 23.0 kPa. The detailed calculation of Figure 4.30 can be found in Appendix F.

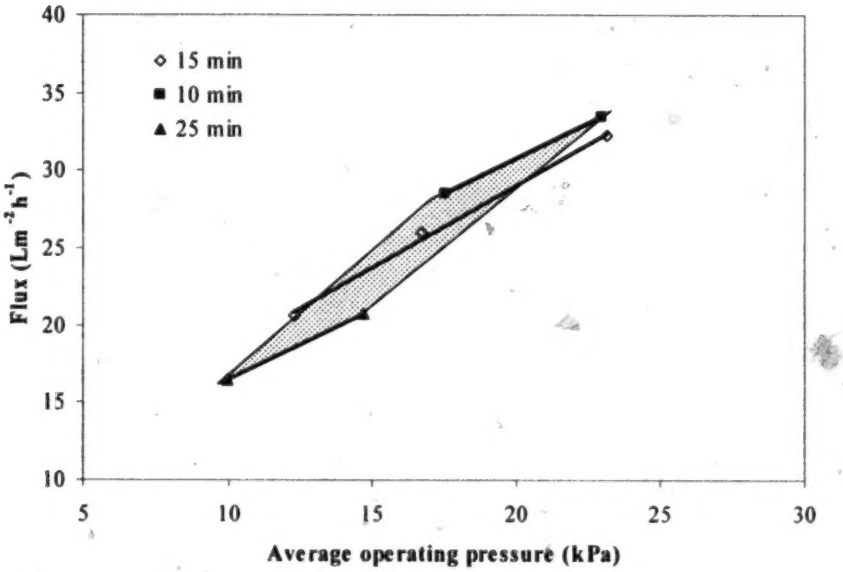

Figure 4.30. Flux versus TMP with different operating time.

4.4 Maintenance and Recovery Cleaning Effectiveness

Maintenance cleaning was conducted when backpulse could not remove the fouling effectively and the membrane system was not achieved good performance, or when the operating parameter set points were changed. The average specific flux in three operating cycles before and after the maintenance cleaning was calculated and listed in Table 4.7 to compare the cleaning effectiveness.

To compare the cleaning effectiveness, the same operating parameter set points before and after the maintenance cleaning should be used. In this project, the operating cycles used to compare the cleaning effectiveness was limited to 10 minutes or 15 minutes operating time; 23.7 $Lm^{-2}h^{-1}$ flux;

237 m^3h^{-1} aeration and 70% water recovery.

By comparison between the NaOCl maintenance cleaning and citric acid maintenance cleaning, the former recovered more specific flux back than the latter one, which suggests that more organic fouling happened in this membrane system. So, it was concluded that NaOCl maintenance cleaning was more significant than the citric acid maintenance cleaning.

When these two maintenance cleanings were taken continuously, but with different sequences, the result showed that the NaOCl cleaning followed by citric acid cleaning recovered more specific flux back. The NaOCl–citric acid cleaning cleaning taken on January 29 and 30, totally recovered the flux of 0.38 $Lm^{-2}h^{-1}kPa^{-1}$, while the citric acid-NaOCl cleaning taken on Febuary12 and the following day resulted in the recovery of only 0.30 $Lm^{-2}h^{-1}kPa^{-1}$. The NaOCl-citric acid cleaning taken on March 22 gained 0.9 $Lm^{-2}h^{-1}kPa^{-1}$ more specific flux back than citric acid-NaOCl cleaning taken on April 8. So NaOCl-citric acid maintenance cleaning sequence is suggested.

When NaOCl-citric acid cleaning was taken on March 22, the citric acid cleaning was taken 2 hours after the NaOCl cleaning, but it still recovered some specific flux back. This illustrates it is necessary to take citric acid maintenance cleaning after NaOCl cleaning.

This ZeeWeed® 500D membrane system began to be operated from January 19, 2004 to October 3, 2004. NaOCl maintenance cleaning was taken once every week. It was found that the cleaning

effectiveness decreased gradually in a long-term, as indicated by the decrease of the specific flux after maintenance cleaning (Figure 4.31). Finally, NaOCl recovery cleaning was taken on October 1. After the recovery cleaning the specific flux was recovered almost back to the specific flux of new membrane. Based on the experience from this project, recovery cleaning once each year will be effective to keep the good membrane performance.

Table 4.7. The comparison of different cleaning efficiency.

	Average specific flux before the cleaning ($Lm^{-2}h^{-1}kPa^{-1}$)	Average specific flux after the cleaning ($Lm^{-2}h^{-1}kPa^{-1}$)	Specific flux ($Lm^{-2}h^{-1}kPa^{-1}$)	Specific flux ($Lm^{-2}h^{-1}kPa^{-1}$) (combined consecutive NaOCl and Citric acid cleaning)	Recovery %
Jan 29 NaOCl [A]	1.40	1.72	0.32		22.4
Jan 30 Citric acid [A]	1.59	1.78	0.19		11.5
				0.38	27.1
Feb 5 NaOCl [A]	1.57	1.82	0.25		15.9
Feb 12 Citric acid [A]	1.64	1.83	0.19		11.2
Feb 13 NaOCl [A]	1.58	1.94	0.36		18.0
				0.3	18.3
Mar 16 NaOCl [A]	1.38	1.72	0.34		24.8
Mar 22 NaOCl [B]	1.37	1.66	0.29		20.1
Mar 22 Citric acid [B]	1.65	1.69	0.04		2.9
				0.32	23.5
Mar 26 NaOCl [B]	1.43	1.68	0.25		17.3
Mar 30 NaOCl [A]	1.40	1.64	0.24		17.5
Apr 8 Citric acid [B]	1.62	1.70	0.08		4.8
Apr 8 NaOCl [B]	1.66	1.85	0.19		10.8
				0.23	14.0

[A] Operating time 15 minutes.

[B] Operating time 10 minutes.

Note: The comparison is based on the same operation situation before and after each cleaning process.

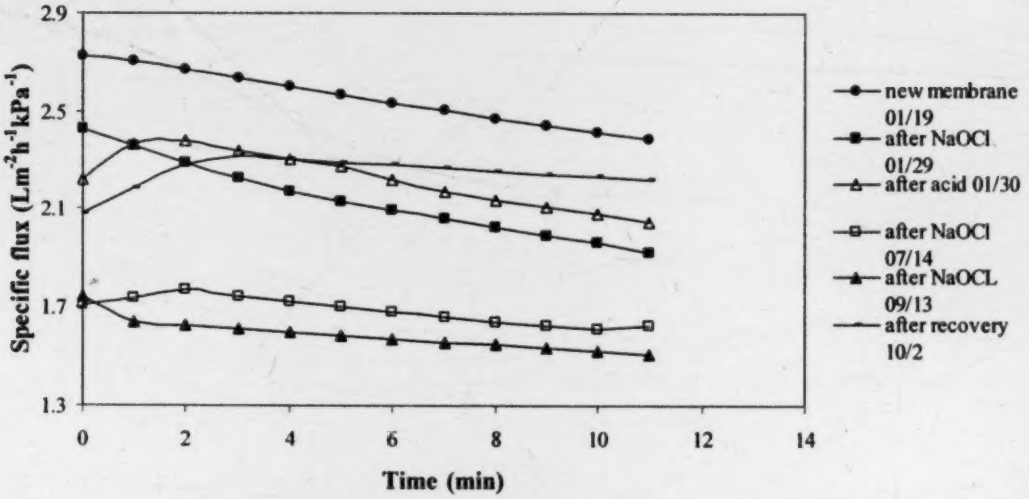

Figure 4.31. Specific flux after different chemical cleaning.

4.5 Membrane Fouling

In this membrane system, one operating cycle includes two parts: production and backpulse. During the production, water is drawn from the membrane tank through the pores in the membrane fibers by a vacuum. Meantime, solids larger than the membrane pores deposit on the membrane surface and form a solid layer even though aeration reduces the deposit. Some of solid layer will be blown-off by the backpulse, named reversible fouling, but small portion of this solid layer cannot be cleaned by backpulse, named irreversible fouling. In this part, the built up of reversible and irreversible fouling is discussed.

The variance of specific flux was used as the indication of fouling. Higher specific flux means

less fouling; lower specific flux means more fouling. The specific flux after one backpulse is always higher than the specific flux before the next backpulse, because more and more solid is building up with the production time. Then an effective backpulse can remove some of these solids or even all of these solids. The left solids become irreversible fouling and decrease the specific flux. The variance of specific flux after backpulse was used to show the extent of the irreversible fouling. The variance of specific flux before backpulse was used to show the extent of the combined reversible and irreversible fouling.

In case of 10 minutes and 15 minutes operating time, flux was 29.4 $Lm^{-2}h^{-1}$ and 23.7 $Lm^{-2}h^{-1}$, respectively. The ZeeWeed® membrane system ran a longer period at these specified conditions. Figure 4.32 and 4.33 show that both the specific flux after backpulse and before backpulse decreased a lot in the first day and then kept relatively stable for many days till the end of the operation.

When the operating time increased to 25 minutes, the ZeeWeed® membrane system did not run longer than 2 days with flux at 23.7 $Lm^{-2}h^{-1}$ and 19.0 $Lm^{-2}h^{-1}$. At the beginning of the production, both of the specific fluxes after backpulse and before backpulse dropped quickly and remained stable for a short period. The specific flux before backpulse with flux at 23.7 $Lm^{-2}h^{-1}$ decreased by 0.4 $Lm^{-2}h^{-1}kPa^{-1}$ from 12 to 14 hours operation and the specific flux before backpulse with flux at 19.0 $Lm^{-2}h^{-1}$ decreased by 0.5 $Lm^{-2}h^{-1}kPa^{-1}$ from 30 hours to 32 hours operation. The changes of specific flux after backpulse for both of them were not very significant. The details of

these changes are presented in Figure 4.34 to Figure 4.38. It was found that the specific flux before backpulse dropped in the two continuous operating cycles suddenly, not gradually. This decrease caused higher TMP required for the same flux, which will squeeze the membrane fouling and make it more difficult to clean the fouling. So it is a better way to do the chemical cleaning before the quick drop of specific flux to achieve a higher chemical cleaning effectiveness.

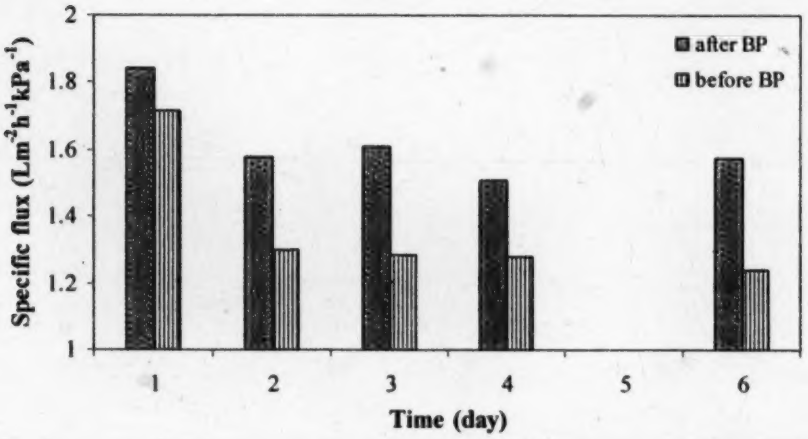

Figure 4.32. Specific flux versus time under 10 minutes operating time, 30 seconds BP, 28.4 $Lm^{-2}h^{-1}$ flux, 237 m^3h^{-1} aeration.

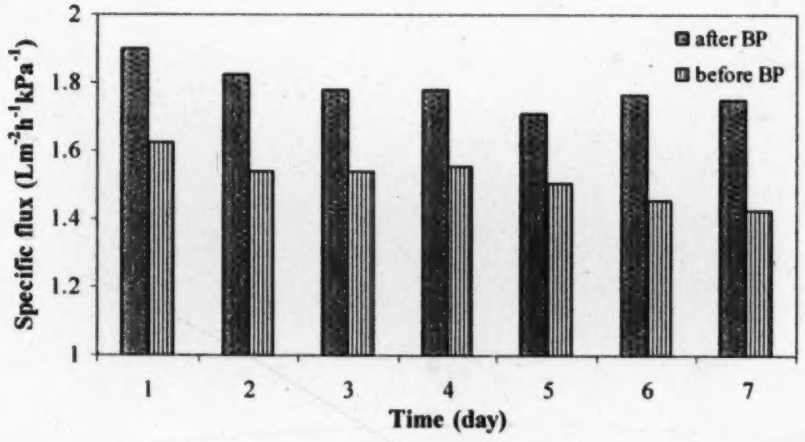

Figure 4.33. Specific flux versus time under 15 minutes operating time, 30 seconds BP, 23.4 $Lm^{-2}h^{-1}$ flux, 237 m^3h^{-1} aeration.

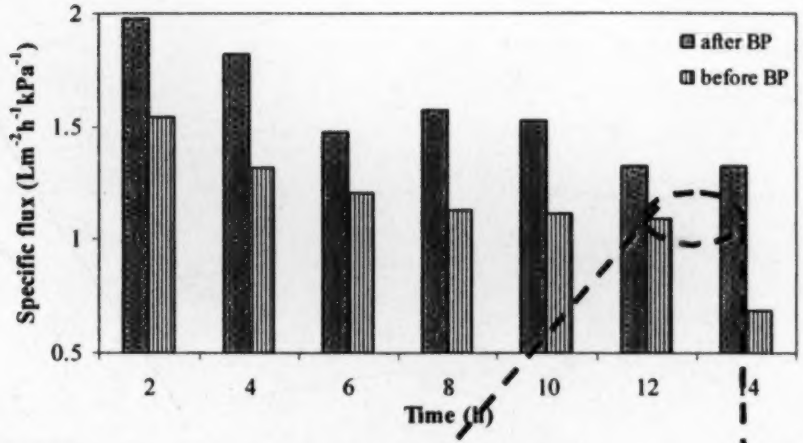

Figure 4.34. Specific flux versus time under 25 minutes operating time, 30 seconds BP, 23.7 $Lm^{-2}h^{-1}$ flux, 237 m^3h^{-1} aeration.

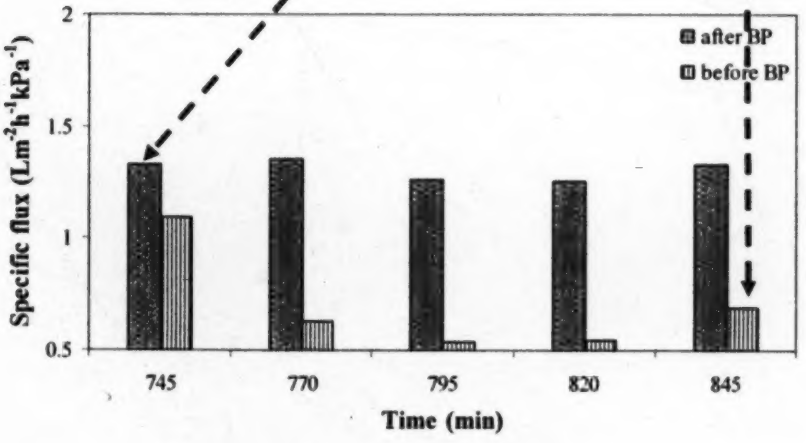

Figure 4.35. Specific flux at certain time.

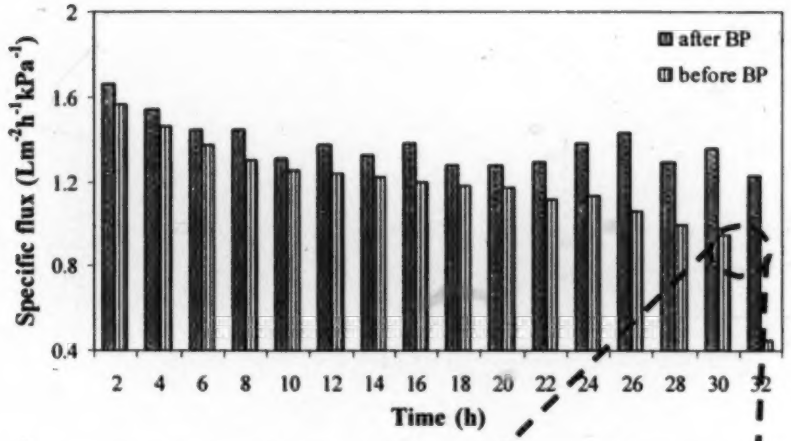

Figure 4.36. Specific flux versus time under 25 minutes operating time, 30 seconds BP, 19.0 $Lm^{-2}h^{-1}$ flux, 237 m^3h^{-1} aeration.

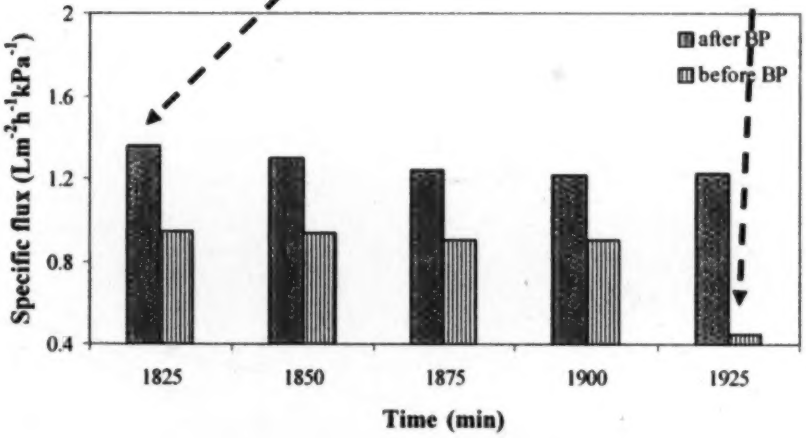

Figure 4.37. Specific flux at certain time.

Figure 4.38 shows the result of specific flux against time after the flux changed from 23.7 $Lm^{-2}h^{-1}$ to 14.3 $Lm^{-2}h^{-1}$. Before the change the membrane irreversible fouling occurred already, as

indicated by the low specific flux at the beginning of this figure. After the change (time 0), both the specific fluxes after and before backpulse were increasing, which means some of the irreversible fouling from the high flux operating was removed by the lower flux operation gradually. Two and half days, the specific flux after backpulse reached a peak point, then decreased due to the new fouling.

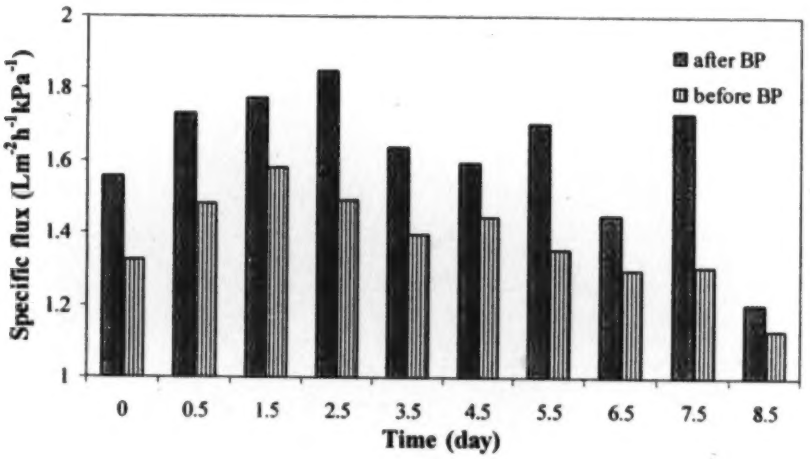

Figure 4.38. Specific flux versus time after a change from high flux to low flux.

4.6 Others

4.6.1 Aeration and Backpulse Cleaning

Except for the chemical cleaning, a temporary pause of production was also tested to recover the fouling after severe fouling occurred. The temporary pause of production, however with aeration on, for several minutes is named as aeration cleaning. Five minutes and ten minutes aeration cleaning were tested to examine the cleaning efficiency. The 5-minute aeration cleaning could not bring the system back from high TMP to low TMP (e.g. Figure 4.39), whereas 10-minute aeration

cleaning did (e.g. Figure 4.40). When the TMP decreases from the higher value to the lower value, it means the membrane system can last for a relatively longer time, but TMP will increase again in a short period.

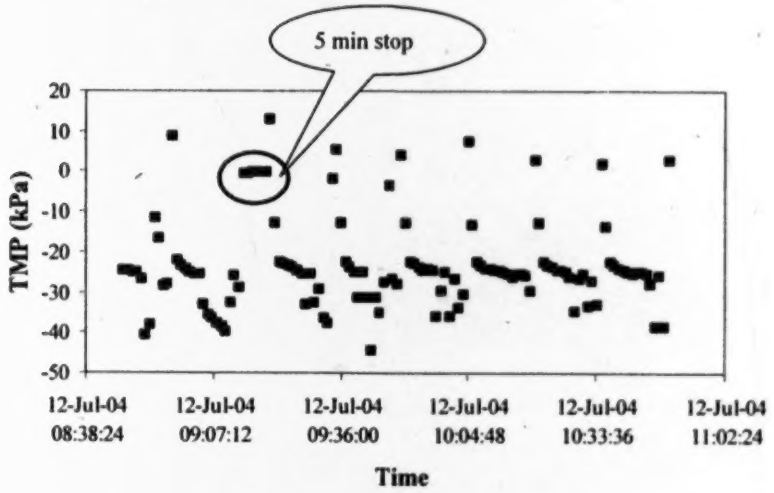

Figure 4.39. TMP before and after 5 minutes shutdown.

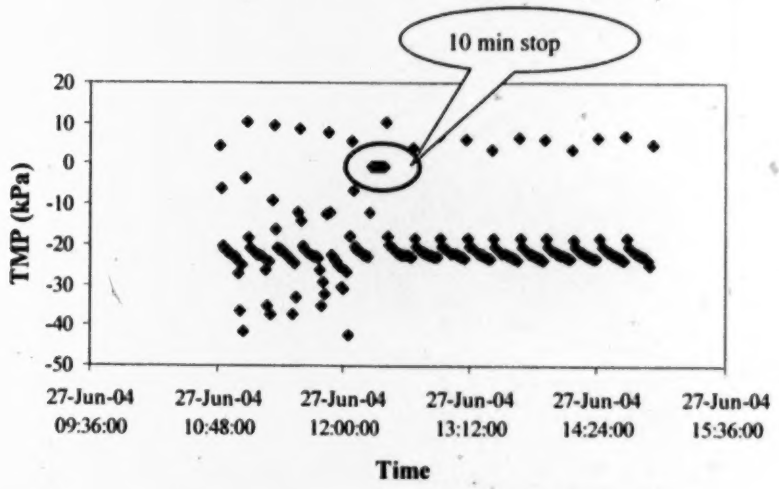

Figure 4.40. TMP before and after 10 minutes shutdown.

4.6.2 Foaming

No foaming was observed during the production process. Very light foaming occurred in the NaOCl maintenance cleaning, except that severe foaming happened during NaOCl cleaning at the end of April and May(Figure 4.41). The foaming is worth of notice and the causes need to be clarified because it causes serious shutdown of this operation unit.

Figure 4.41. Foaming during NaOCl maintenance cleaning.

4.6.3 Energy Input

Power meter was installed for the process pump and blower. These two instruments are the main energy consuming parts in this system. Around 320 kWh was consumed per day and 0.73 kWh was required per cubic meter permeate. The blower consumed most of energy. When the permeate flux was lower, less energy was required, but the change was not very significant.

4.6.4 Membrane Module Appearance

Before the membrane cassette was put to the membrane tank, the color of membrane tubes was pure white. After one-month operation, the color of membrane was light yellow, especially at the top of the membrane (Figure 4.42). Probably some calcium accumulated on the membrane surface. This is because the aeration released from the bottom of the membrane tank could not effectively blow to the top of the membrane and fouling was inclined to accumulate there. After 10 months operation, the membrane cassette was pulled out from the membrane tank. It was found that the color was much darker and some sticky material was accumulated on the membrane surface. Even several membrane tubes were found to stick together (see the Figure 4.43). Clearly if the membrane tubes stick together, the membrane filtration surface will be significantly reduced.

Also lots of snails were found on the membrane surface. Some of them were really small and some of them were bigger (see the Figure 4.44). In this system, a 3 mm strainer was used to pre-filtrate the feed water. It was impossible for the big snail to pass through this strainer. So the snails were actually growing on the membrane surface from a tiny one, or even the snail eggs. If these snails stay on the membrane surface for enough long time, they will grow even much bigger and more and more will grow on the membrane surface and even lead to serious problem for the membrane normal operation. So some solutions should be used to control snail growth and protect the membrane.

Figure 4.42. Membrane appearance after one-month operation.

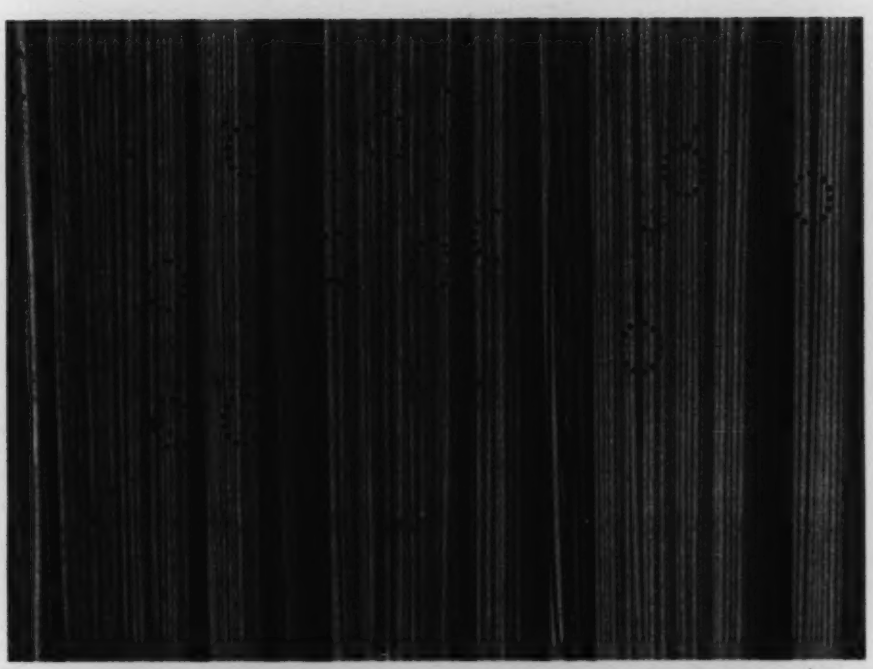

Figure 4.43. Membrane appearance after 10 months operation.

Figure 4.44. Snails taken from the membrane surface.

4.6.5 Utility Water Used as Feed

Before the final effluent from GBWWTP was used as the feed water, utility water was fed to the membrane tank for several days. The operating results show that there was no fouling occurred and the membrane performance did not worsen during that period of time. Figure 4.45 shows the specific flux curves of one operating cycle chosen from December 2, 2003 and January 13, 2004. The membrane system had been operating on utility water for 13 days before January 13. The specific flux was not found decreasing. Also despite the operating time on December 2, 2003 was 30 minutes and the operating time on January 13 was 15 minutes, the average specific flux was 2.5 $Lm^{-2}h^{-1}kPa^{-1}$. Compared with the final effluent used as the feed water, the specific flux was quite high and stable.

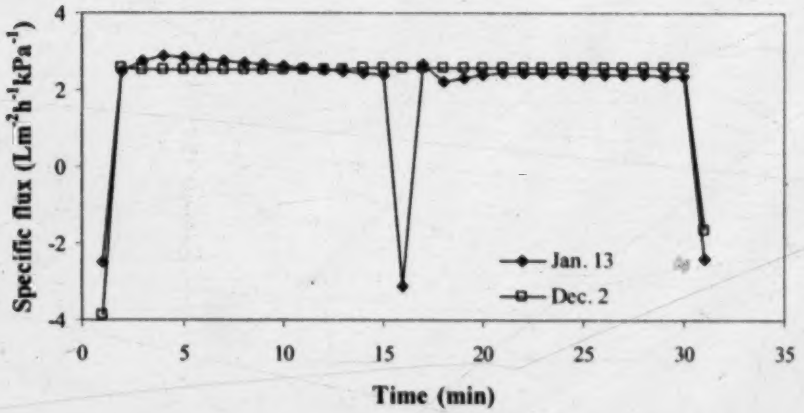

Figure 4.45. TMP against time when utility water used as the feed water.

5 Conclusions

The objective of this study was to evaluate the operation performance of a full size ZeeWeed® 500D membrane system applied as a tertiary wastewater treatment. The membrane system was operated over ten months with two different wastewaters (final effluent after UV and final effluent before UV) as feed water to determine the removal efficiencies of several common wastewater treatment parameters, optimize the membrane operating parameters (flux, operating time, backpulse duration, aeration rate, water recovery) and investigate the chemical cleaning efficiency of both maintenance and recovery cleaning.

On average, permeate contained below 2 mg/L BOD_5, 0.2 NTU turbidity, 0.6 mg/L TSS, and low coliform and coliphage content. Clearly the ultrafiltration membrane exhibited an effective rejection of solid particles and microorganisms of which the sizes were bigger than the nominal pore size of membrane. Nitrogen and phosphorus removals were not very significant. All of above parameters were stable and consistent for ten months of long-term operation regardless of two different feed waters and the membrane operating conditions. Around 30% removal of COD was achieved and permeate COD was influenced by the influent characteristics. In case of metals removal, the effective elimination of 5 metals was achieved by 29%, 17%, 12%, 56%, and 13% for aluminum, cadmium, chromium, copper and lead, respectively.

SDI data were found to be affected by the membrane fouling and operating transmembrane pressure. An increase of operating transmembrane pressure tends to increase SDI. And membrane fouling makes the membrane produce a lower SDI, but also increases the operating pressure that will increase SDI value. Therefore, the increase or decrease of SDI value depends on the competitive effect of membrane fouling and operating transmembrane pressure.

Five membrane operating parameters were investigated in this project including flux, operating time, backpulse duration, aeration amount and water recovery. The optimized backpulse duration for this membrane system was 30 seconds, which could clean the membrane effectively and also keep a relatively high flux. When aeration amount and water recovery were kept in certain range, the variation of flux and operating time had a large impact on the membrane performance. A range of these two parameters was suggested to keep a good membrane performance and to prevent fouling. The membrane flux range was between 16.5 $Lm^{-2}h^{-1}$ and 33.4 $Lm^{-2}h^{-1}$; operating time range was from10 to 25 minutes.

Even though the reversible fouling can be significantly reduced by optimized backpulse, it was impossible to prevent the irreversible fouling completely even in the suggested operating range. For long-term operation, both maintenance and recovery chemical cleaning were necessary. In maintenance cleaning the NaOCl maintenance cleaning is more effective than the citric acid maintenance cleaning, and NaOCl cleaning followed by citric acid cleaning could recover more specific flux back. However the maintenance cleaning effectiveness decreased gradually in a

long-term, so NaOCl recovery cleaning had to be executed. After the recovery cleaning the specific flux was recovered almost back to the specific flux of new membrane.

6 Recommendations

1. NaOCl maintenance cleaning is suggested to be operated once every two weeks and citric acid maintenance cleaning once per month just after the NaOCl maintenance cleaning. NaOCl recovery cleaning probably will be required once a year.
2. The backpulse tank should be constructed separately from the permeate tank, by which it is possible to add chemical to the backpulse tank, sterilize the reverse water and reduce the permeate water bacteria contamination.
3. Attention should be paid to suppress the foaming during NaOCl maintenance cleaning. Otherwise, the foam floating out from the membrane tank will contaminate the working environment and cause health risk for the operators.
4. Snails growing on the membrane surface are concern in the long-term operation. Some special solution should be used, especially in the summer time. Otherwise, the snails will reduce the membrane's life span and increase the membrane replacement cost.
5. The membrane tent should be well ventilated. The aerosol coliform concentration inside of the membrane tent is 10 times higher than the outside, which is a high risk for the operators' health.

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Appendix A: Control Procedure of Standby, Backpulse, Production and Shutdown

Table A.1. Control Procedure of standby, backpulse, production and shutdown

Standby	Step 30	Stop chemical pumps and the process pump and leave valuves as they were. After 10 seconds proceed to the next step.
	Step 31	Hold this step for 30 seconds then proceed to step 32. if standby interlock is cleared, proceed to backpulse step 40, production step 60, or if requested maintenance clean step 100.
	Step 32	Hold this step for 240 seconds then proceed to standby step 33. if standby interlock is cleared, proceed to backpulse step 40, production step 60, or if requested, maintenance clean step 100.
	Step 33	The train will proceed to the backpulse step 40 based on the triggers. Otherwise it will sit in this step indefinitely.
Priming Step	Step 34	Hold this step 5 seconds then proceed to backpulse step 40.
Backpulse	Step 40	Wait until flow through the process pump is low then go to next step.
	Step 41	Set the process pump for backpulsing direction then proceed to the next step.
	Step 42	Start the pump using 60% speed in manual for 3 second then put in auto. Run process pump for backpulse duration set time. If backpulse tank low level is tripped then proceed to next step.
	Step 43	Ramp pump speed down to 0% in 5 seconds, by putting PID loop to manual, then proceed to next step.
	Step 44	Turn off pump. After 1 second delay proceed to production step 60.
Relax	Step 210	Wait for 8 seconds then to step 151.
	Step 211	Hold this step for certain time then go to production step 60.
Production	Step 60	Wait until flow through the process pump is low then go to next step.
	Step 61	Set the process pump pumping direction to produce permeate. Start the pump using 35% in manual for 3 seconds then put it in auto. Go to the next step after the operating time.
	Step 62	Go to the next step.
	Step 63	Go to the next step.
	Step 64	Go to the next step.

	Step 65	At the end of the current production cycle, proceed to backpulse step 40 or relax step 210 or if the "Initiate Full Tank Chlorine Maintenance Clean" or "Initiate Full Tank Acid Maintenance Clean" pushbutton was pressed then proceed to maintenance clean step 100. If the "Initiate Empty Tank Chlorine Maintenance Clean" or "Initiate Empty Tank Acid Maintenance Clean" pushbutton was pressed then proceed maintenance clean step 100.
Shutdown	Step 1	Stop chemical pump and process pump, but not the blower and leave valves as were. After 10 second, proceed to next step. If there is a shutdown alarm, or proceed to step 4 for operator initiated OFF.
	Step 2	Hold this step until level is above membrane then proceed to next step.
	Step 3	Close all valves; proceed to next step (after alarm reset is pressed).
	Step 4	The train will remain OFF until the operator changes the train to another mode of operation.

Appendix B: Maintenance Cleaning Procedure

Table B.1. Normal operation sequence of maintenance cleaning.

Step 100	Aerates for 300 seconds and waits until flow through the process pump is low then go to next step.
Step101	If in empty tank maintenance clean, drain the tank completely, then proceeds to the next step. If in full tank maintenance clean, drains tank to certain level, then proceed to next step.
Step 102	Set the process pump to backpulsing direction in maintenance clean. Go to the next step.
Step 103	Start the process pump using 35% in manual for 3 seconds then put the pump in auto. Hold this step 80 seconds then proceed to the next step. If the backpulse tank low level is tripped then proceed to maintenance clean step 108.
Step 104	Wait 4 seconds then proceed to the next step.
Step 105	Hold this step for 120 seconds then proceed to the next step.
Step 106	Wait 8 seconds then proceed to the next step.
Step 107	Start the process pump using 35% in manual for 3 seconds then put the pump in auton hold this step for 30 seconds & then proceed to maintenance clean step 104. on the 9th iteration or if backpulse tank low level is tripped then proceed to the next step.
Step 108	Wait 4 second then proceed to the next step.
Step 109	Hold this step for 120 seconds then proceed to the next step.
Step 110	Wait 4 second then proceed to the next step.
Step 111	Start the process pump using 35% in manual for 3 seconds then put the pump in auto. Hod this step for 80 seconds. If the backpulse tank low level is tripped, then proceed to the next step.
Step 112	Wait 4 second then proceed to then next step.
Step 113	If the water level is above certain level in empty tank maintenance clean then blower remains off and proceeds to next step. If in full tank maintenance clean starts blower and aerates for 300 seconds then proceeds to next step.
Step 114	Proceed to next step.
Step 115	Hold this step for 20 seconds and until the water level above the membrane, the proceed to Standby step 33.

Appendix C: Recovery Cleaning Procedure

Table C.1. Recovery cleaning procedure.

Soak Step 160	The train will remain in soak until the operator presses initiate chlorine clean or initiate acid clean. The PLC then puts the train to recovery clean step 160.
Step 161	Aerate for 15 minutes then proceed to the next step.
Step 162	Drain the membrane tank to certain level. Set the process pump for backpulse direction in recovery clean then proceed to the next step.
Step 163	Wait 4 seconds then proceed to the next step.
Step 164	Proceed to the next step.
Step 165	Wait 4 seconds then proceed to the next step.
Step 166	Start the process pump with 60% in manual for 3 seconds then put the pump in auto. Hold this step for 80 seconds then proceed to the next step. If backpulse tank water low level is tripped then proceed to recovery clean step 171.
Step 167	Wait 4 seconds then proceed to the next step.
Step 168	Hold this step for 120 seconds then proceed to the next step.
Step 169	Wait 4 seconds then proceed to the next step.
Step 170	Start the process pump using 60% in manual for 3 seconds then put the pump in auto. For 9 iteration hold this step for 30 seconds and then proceed to recovery clean step 167. On the 9th iteration or is backpulse tank low level tripped then go to next step.
Step 171	Wait 4 seconds then proceed to the next step.
Step 172	Provide a banner telling the operator to fill the membrane tank. When the "Tank Filling Complete" Button is pressed and when the level in the tank is above the membrane, then proceed to next step.
Step 173	Proceed to next step.
Step 174	Go to the next step.
Step 175	Hold this step for 3 minutes then proceed to the next step.
Step 176	Hold this step for 6 hours then proceed to the next step. The PLC automatically aerates the tank for 30 seconds every 30 minutes in this step.
Step 177	Hold this step till "Ready to Neutralize" button shows. After this button is pressed then proceed to the next step.
Step 178	Go to the next step.
Step 179	Go to the next step.
Step 180	Hold this step until operator presses the button to "Confirm neutralization" then proceed to next step.
Step 181	Drain the membrane tank completely then proceed to next step.
Step 182	Start the process pump using 60% in manual for 3 seconds then put the pump in auto. Hod this step for 80 seconds. If backpulse tank water low level is tripped then proceed

	to next step.
Step 183	Wait 4 seconds then proceed to the next step.
Step 184	Drain the membrane tank completely then proceed to next step.
Step 185	Wait 10 seconds then proceed to the next step.
Step 186	Proceed to the next step.
Step 187	Proceed to the next step.
Step 188	Proceed to the next step.
Step 189	Operator must restart the feed pump to refill the membrane tank. Hold this step until membrane is submerged then proceed to step 160.

Appendix D: Summary of Operation Record

Table D.1. Summary of all the major operating changes from Dec. 02, 2003 to Sep. 16, 2004.

Date	Time	Operation Changes
12-02-03	16:00	System began to run on the utility water.
12-05-03	15:42	System was shutdown.
12-08-03	08:05	System began to run again.
12-12-03	09:25	Citric acid cleaning
	15:03	Shutdown.
12-15-03	Afternoon	NaOCl cleaning
12-17-03	07:50	Shutdown for service on influent meter.
12-18-03		NaOCl cleaning, after that system was shutdown for holiday.
01-08-04	14:40	Began to run. OP 15 min, BP 30 sec, Per 2.2 L/s
01-16-04	15:57	Shutdown
01-19-04	11:05	Empty membrane tank and change FE after UV as the influent water. OP 15 min, BP 30 sec, Per 5 L/s.
01-22-04	10:00	Influent was throttled.
	11:50	NaOCl cleaning.
01-29-04	13:30	NaOCl cleaning.
01-30-04	12:25	Citric acid cleaning.
02-05-04	14:53	NaOCl cleaning.
02-10-04	04:30	Shutdown
02-11-04		Restart
02-12-04	10:00	Lost data
	13:10	Citric acid cleaning.
02-13-04	11:00	Data began to be recorded.
	13:50	NaOCl cleaning
02-19-04	12:25	Initiated NaOCl cleaning. After that change the set points: OP 10 min, BP 15 sec, Per 5 L/s.
	16:00	Began to lost data.
02-20-04	8:00	Data began to be recorded.
02-26-04	13:10	Initiated NaOCl cleaning.
	13:50	Changed the set points: OP 25 min, BP 30 sec, Per 5 L/s.
02-29-04	18:25	Plant was shutdown due to low permeate flow.
03-01-04		Change set points: OP 25 min, BP 30 sec, Per 3 L/s.
03-02-04	08:48	Began to lost data.
	10:16	Recorded data again.
03-03-04	02:14	Membrane tank low level alarm.

03-09-04	10:25	Initiated NaOCl cleaning. Change set points: OP 25 min, BP 30 sec, Per 4 L/s.
03-09-04	12:40	Changed set points: OP 15m, BP 15 sec, Per 5 L/s.
	21:00	System was shutdown.
03-12-04	10:00	Initiated NaOCl cleaning. Changed set points: OP 15 min, BP 15 sec, Per 4 L/s.
03-16-04	15:48	Changed set points: OP 10 min, BP 30 sec, Per 5 L/s.
03-19-04	11:07	Shutdown the system.
	11:16	Restarted.
	13:54	Began to lost data.
	15:53	Recorded data again.
03-22-04	12:58	Initiated NaOCl cleaning.
	14:42	Citric acid cleaning
	16:30	Changed set points: OP 15 min, BP 15 sec, Per 5 L/s.
03-26-04	7:15	Shutdown due to turbidity alarm. Restarted.
	13:45	Initiated NaOCl cleaning.
	16:15	Changed set points: OP 15 min, BP 30s sec, Per 5 L/s, Air 120 CFM.
03-30-04	10:00	Initiated NaOCl cleaning.
	11:42	Increased air to 140 CFM.
	13:29	Change set points: OP 15 min, BP 30 sec, Per 6 L/s, Air 140 CFM.
03-31-04	14:05	Shutdown the production for 5 min and restarted.
04-02-04	10:00	Initiated NaOCl cleaning.
	11:10	Changed set points: OP 10 min, BP 30 sec, Per 6 L/s, Air 140 CFM.
04-08-04	11:06	Initiated citric acid cleaning.
	13:52	Initiated NaOCl cleaning.
	15:10	Changed set points: OP 10 min, BP 30 sec, Per 6 L/s, Air 120 CFM.
04-14-04	10:34	Initiated NaOCl cleaning.
	12:32	Shutdown for a while and restarted.
04-19-04	9:00	Shutdown for 5 min and restarted.
04-16-04	11:02	Changed set points: OP 10 min, BP 30 sec, Per 5 L/s.
	11:36	Initiated NaOCl cleaning.
04-19-04	09:00	Shutdown for 5 min and restarted.
04-20-04	11:00	Initiated NaOCl cleaning. Changed set points: OP 15 min, BP 15 sec, Per 6 L/s, Air 140 CFM.
04-21-04	06:35	System shutdown due to the low permeate flow.
	08:45	Restarted and changed set points: OP 15 min, BP 30 sec, Per 5 L/s Air 140 CFM.
04-22-04	10:48	Initiated NaOCl cleaning.
	11:28	Changed set points: OP 12.5 min, BP 23 sec, Per 5.5 L/s.
04-26-04	11:25	Initiated NaOCl cleaning.

	12:00	Changed set points: OP 10 min, BP 24 sec, Per 5.3 L/s.
04-30-04	11:17	Initiated NaOCl cleaning.
	12:00	Changed set points: OP 10 min, BP 15 sec, Per 6 L/s, Air 120 CFM.
05-04-04	08:20	Increased air to 140 CFM.
05-04-04	13:30	Initiated NaOCl cleaning.
	14:30	Restarted system and changed set points: OP 25 min, BP 30 sec, Per 3 L/s, Air 120 CFM.
05-14-04	14:30	Restarted the system and changed set points: OP 15 min, BP 30 sec, Per 5 L/s, Air 140 CFM.
05-17-04	02:10	Shutdown due to the high turbidity.
	09:30	Cleaned the permeate tank.
	11:00	Initiated NaOCl cleaning. Added bleach to permeate tank and soaked for a night.
05-18-04	08:55	Drained the permeate tank and restarted the system.
05-19-04	08:16	Power failure alarm.
05-20-04	11:30	Shutdown the system for maintenance. Change the strainer from 1/8 (3 mm) inch to 1/32 (1 mm) inch.
	15:00	Initiated citric acid cleaning.
	16:00	Initiated NaOCl cleaning. Set points: OP 15 min, BP 30 sec, Per 5 L/s, Air 140 CFM.
05-25-04	09:35	Initiated NaOCl cleaning.
	10:14	Back to the normal operation.
05-26-04	13:30	Shutdown the system and changed the strainer back.
05-28-04	08:45	Changed the FE from tank 6 as the influent water. Set points: OP25 min, BP 30 sec, Per 3 L/s, Air 140 CFM due to low influent flow rate.
06-02-04	08:30	Changed the influent pump. Set points: OP 15 min, BP 30 sec, Per 5 L/s, Air 140 CFM.
06-07-04	10:49	Initiated NaOCl cleaning.
	11:30	Sept points: OP 10 min, BP 30 sec, Per 5.5 L/s.
06-11-04	11:00	Initiated NaOCl cleaning.
	12:40	Changed set points: OP 15 min, BP 30 sec, Per 5 L/s.
06-19-04		Shut down due to high turbidity.
06-22-04	09:00	Restarted.
06-23-04	08:00	Shut down to install RO system.
06-27-04	12:05	Shut down for 10 min.
06-28-04	09:50	Shutdown the system and did 50 sec backpulse manually, restarted.
	11:25	Initiated NaOCl cleaning.
	13:37	Changed set points: OP 15 min, BP 40 sec, Per 5 L/s, Air 140 CFM.
06-30-04	08:45	Initiated citric acid cleaning.

07-05-06	09:40	Shut down the system due to the permeate pump leaking.
07-12-04	08:50	Shut down for 5 min.
	13:50	Shut down for 10 min.
	15:05	Finished back pulse for 2 min manually and restarted.
07-14-04	09:25	Initiated NaOCl cleaning.
	11:00	Changed set points: OP 15 min, BP 30 sec, Per 5 L/s.
07-17-04	11:50	Shut down the permeate production and cleaned the turbidity meter.
07-21-04	10:10	Shut down the permeate production and cleaned the turbidity meter.
07-27-04	8:45	Shut down for 5 min and restarted and set the permeate at 4 L/s.
	13:00	Initiated NaOCl cleaning.
		Upgrade the PLC software.
07-28-04		Changed the permeate back to 5 L/s.
07-30-04	09:10	Initiated NaOCl cleaning. Changed the permeate 3 L/s, 4 L/s and 5 L/s for SDI test.
	14:00	Changed the set points: OP 15 min, BP 30 sec, Per 5 L/s, Rec 90 %.
08-11-04	09:10	Initiated NaOCl cleaning.
08-12-04	12:50	Shut down for 5 min to install the power meter back.
08-18-04	05:06	Both aeration valves keep open.
08-20-04	09:20	Initiated NaOCl cleaning. One of the water level switch is out of control.
08-23-04	08:50	Initiated NaOCl cleaning.
08-24-04	08:40	Changed the water recovery from 90 % to 95 %.
08-31-04	09:53	Changed the permeate set point from 5 L/s to 4.5 L/s.
09-02-04	13:00	Shut down the system to clean influent pump.
	13:30	Initiated NaOCl cleaning.
	15:07	Changed the set points: OP 15 min, BP 30 sec, Per 5 L/s, Air 150 CFM.
09-13-04	09:25	Initiated NaOCl cleaning.
09-15-04	10:30	Initiated citric acid cleaning.
09-16-04		Changed permeate set point at 3 L/s and water recovery at 80 %.

Appendix E: Water Quality Raw Data

Table E.1. Raw data for June after change the FE before UV as the feed water.

			28-May	31-May	2-Jun	4-Jun	7-Jun	9-Jun	11-Jun	18-Jun	21-Jun	23-Jun	28-Jun	30-Jun
Influent	BOD-C	mg/L	~10	3.4	2.6	~7.1	2.4	<2	<2	~2.1	8.5	~7.1	~3.9	2.5
	CO-F-M	CFU/100mL	~7500	18400	~8600	~9400	3600	4500	~8000	22000	22000	170000	~6800	5300
	CO-T-M	CFU/100mL	300000	280000	200000	130000	~76000	79000	79000	~164000	15000	1500000	~105000	~112000
	COD	mg/L	47	32	34	36	37	38	36	32	27	55	39	40
	$N\text{-}NH_3$	mg/L	0.667	0.803	0.0400	0.0210	0.0200	0.00400	0.0220	0.0330	1.45	10.3	0.206	0.0170
	N-TKN	mg/L	3.05	2.28	1.83	1.70	1.67	1.69	1.60	1.81	3.24	10.5	2.07	1.83
	N-TOX	mg/L	7.78	7.79	5.26	4.84	7.68	5.80	3.94	0.249	0.476	1.89	6.16	4.29
	TP	mg/L	0.968	0.345	0.319	0.265	0.327	1.07	0.430	0.454	0.200	1.99	0.441	0.474
	TSS	mg/L	12.0	5.0	5.4	~3.1	3.9	3.8	5.2	5.9	2.2	6.8	4.9	5.9
Membrane Tank	CO-F-M	CFU/100mL	20800	31200	22000	19000	~6500	~11100	33000	33000	25000	490000	~6300	16000
	CO-T-M	CFU/100mL	500000	530000	380000	320000	~107000	~141000	350000	440000	220000	6300000	~143000	~130000
	TSS	mg/L	23.0	22.0	21.0	12.0	7.5	12.0	10.0	29.0	4.1	18.0	17.0	16.0
Permeate	BOD-C	mg/L	<2	<2	<2	<2	<2	<2	<2	<2	<2	<2	<2	
	CO-F-M	mg/L	<1	<1	<1	2	6	<1	<1	<1	<1	<1	5	<1
	CO-T-M	mg/L	~88	<1	<1	43	~240	<1	10	<1	1	4	~85	2
	$CO_3^=$	mg/L	<1.00	<1.00	<1.00	<1.00	<1.00	<1.00	<1.00	<1.00	<1.00	<1.00	<1.00	<1.00
	COD	mg/L	28	24	25	26	26	25	32	17	21	35	29	74
	HCO_3^-	mg/L	185	181	181	199	151	213	169	198	169	236	163	185
	$N\text{-}NH_3$	mg/L	0.858	0.726	0.0400	0.0240	0.0260	0.00400	0.0300	0.0350	1.93	10.8	0.172	0.0290
	N-TKN	mg/L	1.99	1.64	1.06	1.14	1.06	1.05	1.12	1.04	2.88	10.0	1.16	1.11
	N-TOX	mg/L	7.52	8.16	5.52	5.10	7.49	5.90	4.17	0.239	9.30	2.16	6.51	4.10
	TA	mg/L	185	181	181	199	151	213	169	198	169	236	163	185
	TH	mg/L	273	288	288	297	288	297	236	271	252	258	255	255
	TP	mg/L	0.387	0.0980	0.0750	0.126	0.0720	0.909	0.258	0.0990	0.166	2.02	0.236	0.207
	TSS	mg/L	1.0	<0.6	<0.6	<0.6	<0.6	<0.6	<0.6	<0.6	<0.6	<0.6	<0.6	<0.6

Table E.2. Raw data for July water quality.

			5-Jul	7-Jul	9-Jul	12-Jul	14-Jul	16-Jul	19-Jul	21-Jul	23-Jul	26-Jul	28-Jul	30-Jul
Influent	BOD-C	mg/L	<2	<2	<2	2.4	2.6	<2	<2.0		<2	2.8	14	<2
	CO-F-M	CFU/100mL	13000	3600	5500	2800	7200	6700	5400		~8500	4900	NR1	~15000
	CO-T-M	CFU/100mL	230000	~101000	~101000	78000	140000	240000	29000		~108000	~84000	NR1	410000
	COD	mg/L	40	30	28	32	37	33	33		32	38	61	21
	$N\text{-}NH_3$	mg/L	0.289	0.118	0.0180	0.0340	0.00800	0.00800	0.0370		0.0320	0.0410		2.52
	N-TKN	mg/L	1.75	1.45	1.33	1.53	1.60	1.51	1.40		1.44	1.79	17.8	4.28
	N-TOX	mg/L	6.90	3.24	3.46	11.4	4.77	4.88	7.88		7.46	4.96	2.55	~0.151
	TP	mg/L	0.310	0.422	0.207	0.295	0.445	0.554	1.79		0.227	0.337	4.61	1.67
	TSS	mg/L	5.7	4.3	5.0	7.0	6.2	3.2	2.8		2.5	5.6	8.4	0.8
Membrane Tank	CO-F-M	CFU/100mL	32000	4600	5600	7000	20000	30000	~14700		18000	16000	NR1	
	CO-T-M	CFU/100mL	370000	180000	230000	200000	190000	390000	290000		280000	320000	NR1	
	TSS	mg/L	21.0	15.0	11.0	22.0	24.0	13.0	14.0		11.0	17.0	41.0	
Permeate	BOD-C	mg/L	<2	<2	<2	<2	<2	<2	<2.0	<2	<2	<2	2.9	3.6
	CO-F-M	mg/L	<1	<1	<1	5	7	11	<1	<1	<1	<1	NR1	<1
	CO-T-M	mg/L	<1	1	2	160	190	220	16	42	2	25	NR1	3
	CO_3=	mg/L	<1.00	<1.00	<1.00	<1.00	<1.00	<1.00	<1.00	<1.00	<1.00	<1.00	<1.00	<1.00
	COD	mg/L	26	19	21	21	24	24	24	9	25	28	30	37
	HCO_3-	mg/L	194	170	194	180	228	226	176	210	184	180	259	191
	$N\text{-}NH_3$	mg/L	0.317	0.0620	0.0220	0.0740	0.0300	0.0250	0.0500	0.0320	0.0390	0.0570		2.18
	N-TKN	mg/L	1.14	0.778	0.770	0.949	1.11	1.08	0.935	1.05	1.02	1.11	15.8	4.71
	N-TOX	mg/L	6.74	2.90	3.33	12.1	4.92	5.14	8.20	6.28	7.19	4.82	2.52	~0.168
	?A	mg/L	194	170	194	180	228	226	176	210	184	180	259	191
	TH	mg/L	406	312	~361	453	462	408	346	368	330	318	330	300
	TP	mg/L	0.108	0.262	0.0820	0.0890	0.333	0.555	1.83	0.457	0.132	0.142	4.22	1.50
	TSS	mg/L	<0.6	1.2	<0.6	<0.6	0.6	<0.6	<0.6	<0.6	<0.6	0.7	<0.6	5.8

Table E.3. Raw data for August water quality.

			4-Aug	6-Aug	9-Aug	11-Aug	13-Aug	16-Aug	18-Aug	23-Aug	25-Aug	27-Aug	30-Aug
Influent	BOD-C	mg/L	<2	2.1	<2	<2	<2	<2	<2	<2	2.5	<2	~5.4
	CO-F-M	CFU/100mL	~6900	25000	30000	~8300	13000	26000	7000	~6400	1100	81000	18000
	CO-T-M	CFU/100mL	80000	340000	300000	250000	310000	480000	160000	180000	79000	120000	97000
	COD	mg/L	33	32	29	27	29	29	24	39	30	26	33
	$N\text{-}NH_3$	mg/L	0.0430	0.0830	0.0630	0.0300	0.0620	0.0810	0.0290	0.0310	0.0580	0.0240	0.0630
	N-TKN	mg/L	1.70	1.39	1.42	1.53	1.44	1.33	1.22	1.76	1.31	1.40	1.75
	N-TOX	mg/L	4.75	2.09	10.8	4.92	7.11	8.44	3.44	15.6	4.81	4.19	4.95
	TP	mg/L	0.479	0.370		1.19	0.321	0.323	0.317	4.18	0.829	0.417	0.480
	TSS	mg/L	3.6	6.6	4.3	2.9	2.3	3.0	2.2	13.0	1.2	2.40	3.50
Membrane Tank	CO-F-M	CFU/100mL	20000	45000	~63000	16000	29000	60000	23000	5500	58000	70000	87000
	CO-T-M	CFU/100mL	490000	1200000	~880000	400000	1700000	930000	NR1	210000	1000000	1200000	5300000
	TSS	mg/L	53.0	45.0	27.0	14.0	43.0	41.0	21.0	32.0	14.0	23.0	54.0
Permeate	BOD-C	mg/L	<2	<2	<2	<2	<2	<2	<2	<2	<2	<2	<2
	CO-F-M	mg/L	<1	<1	<1	2	<1	<1	8	<1	<1	<1	<1
	CO-T-M	mg/L	170	140	710000	290	<1	<1	~330	<1	1	2	<1
	CO_3=	mg/L	<1.00	<1.00	<1.00	<1.00	<1.00	<1.00	<1.00	<1.00	<1.00	<1.00	<1.00
	COD	mg/L	21	20	16	21	17	21	20	19	23	22	23
	HCO_3-	mg/L	197	171	173	200	190	203	190	122	194	199	181
	$N\text{-}NH_3$	mg/L	0.0650	0.0570	0.0810	0.0330	0.0750	0.0960	0.0520	0.0430	0.0640	0.0380	0.0940
	N-TKN	mg/L	1.11	0.938	1.05	1.02	1.04	0.985	0.907	0.865	0.920	1.13	1.18
	N-TOX	mg/L	5.09	2.06	11.3	4.80	7.07	8.96	3.94	8.40	5.08	4.01	5.23
	TA	mg/L	197	171	173	200	190	203	190	122	194	199	181
	TH	mg/L	318	nr4	309	306	291	300	297	222	306	306	294
	TP	mg/L	0.242	0.141		1.08	0.232	0.210	0.184	3.63	0.796	0.633	0.331
	TSS	mg/L	<0.6	1.6	<0.6	<0.6	<0.6	<0.6	<0.6	<0.6	<0.6	<0.600	<0.600

Table E.4. Raw data for September water quality.

			8-Sep	10-Sep	13-Sep
Influent after screen	BOD-C	mg/L	~2.4	<2	2.2
	CO-F-M	CFU/100mL	~7300	10000	13000
	CO-T-M	CFU/100mL	1100000	180000	67000
	COD	mg/L	39	29	29
	$N\text{-}NH_3$	mg/L	0.0600	0.0270	0.0950
	N-TKN	mg/L	1.67	1.08	1.46
	N-TOX	mg/L	6.07	3.53	5.71
	TP	mg/L	0.649	0.238	0.311
	TSS	mg/L	3.70	~1.8	6.80
Membrane Tank	CO-F-M	CFU/100mL	26000	43000	21000
	CO-T-M	CFU/100mL	220000	~990000	450000
	TSS	mg/L	39.0	33.3	20.0
Permeate	BOD-C	mg/L	<2	<2	<2
	CO-F-M	mg/L	<1	<1	5
	CO-T-M	mg/L	22	15	180
	CO_3^{2-}	mg/L	<1.00	<1.00	<1.00
	COD	mg/L	27	21	19
	HCO_3^-	mg/L	187	134	194
	$N\text{-}NH_3$	mg/L	0.0800	0.0450	0.111
	N-TKN	mg/L	1.13	0.758	0.989
	N-TOX	mg/L	6.75	3.61	6.05
	TA	mg/L	187	134	194
	TH	mg/L	302	233	324
	TP	mg/L	0.621	0.134	0.218
	TSS	mg/L	<0.600	<0.600	<0.600

Table E.5. January to May part of raw data.

	PERMEATE						INFLUENT					
Date	BOD-C	COD	TP	TKN	NH3-N	TOX-N	BOD-C	COD	TP	TKN	NH3-N	TOX-N
20040119	mg/L	mg/L	mg/L	Mg/L	mg/L	mg/L	mg/L	mg/L	mg/L	mg/L	mg/L	mg/L
20040121	<2.0	24	0.496	6.988	4.218	12.437	4.5	52	1.264	8.961	4.461	12.172
20040123	<2.0	28	0.777	11.981	9.263	11.764	2.5	42	0.711	12.30	9.369	11.711
20040126	<2.0	23	0.192	7.572	7.044	11.167	2.5	35	0.385	8.175	6.418	11.483
20040128	<2.0	26	1.051	16.641	14.371	9.851	2.5	41	1.236	17.338	14.774	9.932
20040130	<2.0	26	0.464	15.537	12.197	11.115	2.8	38	0.769	16.102	12.229	12.31
20040202	<2.0	22	0.487	18.515	13.446	9.095	2.4	37	0.670	18.316	14.616	9.09
20040204	<2.0	20	0.264	16.251	13.170	9.427	2.2	27	0.453	16.485	11.783	9.519
20040206	<2.0	21	0.243	12.664	8.636	9.214	2.3	38	0.401	13.169	9.609	9.384
20040209	2.1	21	0.171	11.436	10.119	10.852	3.8	28	0.428	12.069	9.611	10.583
20040211	<2.0	17	0.164	6.095	5.216	8.997	3.4	33	nr4	6.470	4.416	8.919
20040213	<2.0	23	0.300	5.079	3.583	7.944	~2.9	35	0.331	5.403	3.788	8.118
20040217	<2.0	25	0.778	4.723	2.223	9.573	~2.5	35	1.011	5.313	2.162	9.524
20040218	<2.0	16	0.147	3.920	3.101	9.144	3.2	34	0.415	4.393	2.900	9.174
20040220	<2.0	23	0.205	5.346	4.691	8.207	12	47	0.516	5.623	nr4	9.573
20040223	<2.0	26	0.197	5.905	5.040	6.631	7.9	33	0.305	6.763	5.263	6.553
20040225	<2.0	20	0.26	9.355	8.125	8.476	3.4	36	0.338	10.466	8.005	8.276
20040227	<2.0		0.154	12.489	9.632	11.336	4.6	43	0.605	14.002	8.471	11.098
20040301	nr4	20	1.24	12.1	10.2	8.15	~2.4	34	1.99	11.9	9.44	8.32
20040303	nr4	25	0.374	14.7	17.5	11.0	3	34	0.761	15.0	12.3	11.0
20040305	<2	21	0.528	20.7	14.2	11.1	5	43	0.916	20.2	12.1	10.8
20040310	<2	21	0.509	20.3	17.2	9.53	13	38	0.640	20.5	16.5	9.65
20040312	<2	20	0.652	6.39	5.71	9.34	3	33	1.12	7.00	5.62	9.40
20040315	<2	25	0.411	7.48	6.47	8.49	4	41	0.649	8.89	6.74	8.32
20040317	<2	25	0.150	7.45	6.51	10.2	3	34	0.538	7.99	6.42	10.2
20040319	<2	28	0.105	10.9	8.73	9.15	3	39	0.267	11.8	8.64	8.90
20040322	<2	24	0.144	8.91	8.26	11.2	2	37	0.339	9.72	8.38	11.0
20040324	<2	35	0.869	8.53	7.98	11.8	2	24	0.995	9.15	7.96	11.4

20040331	<2	20	0.729	7.09	5.49	10.6	<2	35	0.928	7.63	5.87	10.6
20040402	<2	28	1.04	9.93	7.83	8.87	3	41	1.16	9.78	7.71	8.46
20040405	<2	54	0.518	8.36	7.64	12.6	3	nr4	0.790	8.77	7.48	14.0
20040414	<2	22	0.848	9.68	7.76	13.0	3	34	0.958	10.8	8.31	12.4
20040416	2	29	0.701	6.51	5.74	14.1	4	44	0.817	6.84	5.34	12.7
20040419	<2	27	0.280	3.27	2.14	11.8	2	27	0.359	3.68	2.23	11.8
20040421	<2	25	1.04	3.11	1.92	14.3	nr4	37	0.984	3.60	1.98	14.1
20040423	<2	24	0.531	4.21	3.32	13.0	2	40	0.594	4.72	3.43	12.7
20040426	<2	29	0.143	1.85	0.736	16.4	4	39	0.320	2.75	0.750	15.8
20040428	<2	28	0.187	1.49	0.545	15.9	3	33	0.314	1.94	0.567	15.1
20040430	<2	32	0.227	1.35	0.307	5.31	<2	44	0.498	2.00	0.369	4.93
20040503	<2	25	0.242	2.72	1.58	6.75	2	32	0.342	3.37	1.59	6.60
20040505	<2	24	0.224	4.72	3.38	9.25	3	31	0.405	5.13	3.37	8.39
20040507	<2	22	1.17	4.62	3.39	5.85	3	38	1.31	5.33	3.20	5.42
20040510	<2	23	0.403	3.51	3.51	6.67	<2	30	0.497	4.55	1.20	5.76
20040518	<2	24	0.154	4.23	3.27	8.55	3	39	0.369	5.21	3.42	8.11
20040519	<2	23	nr5	3.70	2.48	5.56	4		0.477	4.54	2.52	5.54
20040521	<2	24	0.329	8.52	6.86	6.15	4	39	0.484	9.19	6.46	5.72
	<2	26	0.423	7.93	6.24	7.87	4	40	0.632	9.20	6.47	8.58

Table E.6. January to May coliform raw data.

	INFLUENT		PERMEATE		MEMBRANE TANK	
DATE	TC (CFU/100mL)	FC (CFU/100mL)	TC (CFU/100mL)	FC (CFU/100mL)	TC (CFU/100mL)	FC (CFU/100mL)
20040119	700	59	<1	<1		
20040121	810	45	<1	<1		
20040123	790	11	<1	<1		
20040126	310	17	<1	<1		
20040128	740	100	1	<1		
20040130	1600	120	1	<1		
20040202	660	52	1	<1		
20040204	750	12	<1	<1	5800	98
20040206	1600	42	8	<1	12900	140
20040209	2700	88	<1	<1	TNTC	130
20040211	790	24	1	<1	6700	130
20040213	720	26	<1	<1	~8500	100
20040217	2400	17	18	<1	41000	120
20040218	~1040	50	<1	<1	38000	210
20040220	TNTC	100	<1	<1	20000	180
20040223	~1040	27	47	<1	39000	200
20040225	~1160	38	<1	<1	36000	260
20040227	600	13	<1	<1	17000	68
20040301	~6300	38	3	<1	43000	130
20040303	2520	150	<1	<1	47700	230
20040305	3060	190	4	<1	44100	480
20040310	~1280	~97	<1	<1	35100	370
20040312	850	20	<1	<1	62100	320
20040315	11000	70	1300	<1	3500000	230
20040317	1800	~73	<1	<1	~9900	210
20040319	2520	60	14	<1	~8280	140

20040322	2160	18	3	<1	~7380	180
20040324	7200	100	<1	<1	18900	230
20040326	<1	<1	3	<1	4680	110
20040329	1980	29	1	<1	~9270	140
20040331	1890	120	14	<1	~7830	420
20040402	3800	38	25	<1		
20040405	4500	33	42	<1	7800	140
20040414	TNTC	TNTC	<1	<1	46000	3500
20040416	600	23	140	2	12000	190
20040419	3100	48	92	<1	~10200	170
20040421	2700	60	2	<1	6700	130
20040423	2000	35	6	<1	6900	370
20040426	520	29	<1	<1	4000	330
20040428	460	51	29	<1	6900	240
20040430	~1060	35	<1	<1		
20040503	3000	31	140	<1	~9000	160
20040505	3200	96	12	<1	~9200	200
20040507	3500	36	270	<1	4700	120
20040510	790	42	6	<1	2700	100
20040514	235	8	36	<1	5100	44
20040518	~8300	~71	<1	<1	29000	80
20040519	TNTC	86	13	<1	68000	120
20040521	5200	220	<1	<1	28000	560

Table E.7. Data Summary for Metal Analysis*.

Iron name	Average reduction[A] (%)	Average concentrate[B] (%)	May 5th (µg/L)			May 7th (µg/L)			May 10th (µg/L)		
			Influent	Overflow	Permeate	Influent	Overflow	Permeate	Influent	Overflow	Permeate
Aluminum	29	161	0.037	0.075	0.023	0.025	0.036	0.020	0.030	0.041	0.021
Aluminum-d			0.020	0.016	0.015	nr5	nr5	nr5	0.019	0.018	0.015
Antimony	11	100	0.0004	0.0004	0.0004	0.0005	0.0005	0.0005	0.0003	0.0003	0.0002
Antimony-d			0.0002	0.0004	0.0004	nr5	n5	nr5	0.0002	0.0002	0.0002
Arsenic		110	0.0007	0.0008	0.0007	0.0007	0.0008	0.0008	0.0006	0.0006	0.0006
Arsenic-d			0.0006	0.0006	0.0008	nr5	nr5	nr5	0.0006	0.0006	0.0006
Barium	8	99	0.027	0.028	0.025	0.031	0.031	0.029	0.033	0.031	0.030
Beryllium			<0.0001	<0.0001	<0.0001	<0.0001	<0.0001	<0.0001	<0.0001	<0.0001	<0.0001
Beryllium-d			<0.0001	<0.0001	<0.0001	nr5	nr5	nr5	<0.0001	<0.0001	<0.0001
Bismuth			<0.0005	<0.0005	<0.0005	<0.0005	<0.0005	<0.0005	<0.0005	<0.0005	<0.0005
Bismuth-d			<0.0005	<0.0005	<0.0005	nr5	nr5	nr5	<0.0005	<0.0005	<0.0005
Boron		102	0.213	0.226	226	0.204	0.208	0.212	0.228	0.223	0.234
Boron-d			0.215	0.205	0.195	nr5	nr5	nr5	0.224	0.220	0.226
Cadmium	17	125	0.00004	0.00006	0.00003	0.00008	0.00009	0.00007	0.00008	0.00009	0.00007
Cadmium-d			0.00009	0.00014	0.00010	nr5	nr5	nr5	0.00008	0.00008	0.00007
Calcium			63.9	65.5	63.5	66.2	65.0	65.4	65.0	61.6	66.0
Calcium-d			64.8	60.6	63.9	nr5	nr5	nr5	64.6	65.4	64.9
Chromium	12	114	0.0018	0.0021	0.0016	0.0015	0.0017	0.0013	0.0009	0.0010	0.0008

Chromium 6+			nr5	nr5	nr5	nr5	nr5	nr5	nr5	nr5	nr5
Chromium-d			0.0009	0.0009	0.0014	nr5	nr5	nr5	0.0009	0.0008	0.0008
Cobalt		117	0.0007	0.0008	0.0007	0.0008	0.0009	0.0008	0.0004	0.0005	0.0004
Cobalt-d			0.0005	0.0006	0.0006	nr5	nr5	nr5	0.0004	0.0004	0.0004
Copper	56	156	0.003	0.005	0.001	0.003	0.004	0.001	0.003	0.005	0.002
Copper-d			0.004	0.006	0.002	nr5	nr5	nr5	0.003	0.005	<0.001
Iron			<0.1	0.2	<0.1	<0.1	0.1	<0.1	<0.1	<0.1	<0.1
Iron-d			0.06	0.06	0.07	nr5	nr5	nr5	0.06	0.06	0.05
Lead	13	113	0.0011	0.0013	0.0009	0.0012	0.0013	0.0011	0.0009	0.0010	0.0008
Lead-d			0.0008	0.0009	0.0009	nr5	nr5	nr5	0.0008	0.0008	0.0007
Lithium		100	0.028	0.028	0.029	0.030	0.030	0.030	0.028	0.028	0.030
Lithium-d			0.026	0.024	0.028	nr5	nr5	nr5	0.027	0.027	0.028
Magnesium		99	28.7	29.3	28.6	31.6	31.0	31.1	33.3	32.2	34.0
Magnesium-d			32.6	26.4	29.6	nr5	nr5	nr5	32.3	32.6	32.5
Manganese	5	109	0.078	0.088	0.074	0.065	0.070	0.063	0.053	0.057	0.049
Manganese-d			0.062	0.026	0.077	nr5	nr5	nr5	0.049	0.028	0.048
Mercury			nr5	nr5	nr5	nr5	nr5	nr5	nr5	nr5	nr5
Mercury-d			nr5	nr5	nr5	nr5	nr5	nr5	nr5	nr5	nr5
Molybdenum		99	0.020	0.021	0.020	0.015	0.014	0.015	0.007	0.007	0.007
Molybdenum-d			0.006	0.009	0.020	nr5	nr5	nr5	0.006	0.007	0.007
Nickel	8	97	0.0119	0.0121	0.0112	0.0119	0.0121	0.0117	0.0071	0.0062	0.0060
Nickel-d			0.0047	0.0082	0.0108	nr5	nr5	nr5	0.0043	0.0045	0.0042
Potassium	3	99	12.4	12.6	12.0	14.4	14.0	13.8	12.8	12.4	12.7

Potassium-d			12.1	11.9	12.4	nr5	nr5	nr5	12.2	12.2	12.1
Selenium			0.0006	0.0005	0.0004	0.0005	0.0005	0.0006	<0.0002	0.0005	0.0006
Selenium-d			0.0005	0.0003	0.0006	nr5	nr5	nr5	0.0002	0.0005	0.0005
Silicon	1	99	3.68	3.76	3.61	3.93	3.80	3.81	3.72	3.62	3.78
Silicon-d			3.95	3.89	3.90	nr5	nr5	nr5	3.87	3.88	3.90
Silver		233	0.0002	0.0004	<0.0001	0.0001	0.0003	<0.0001	0.0001	0.0002	<0.0001
Silver-d			<0.0001	0.0001	<0.0001	nr5	nr5	nr5	<0.0001	0.0001	<0.0001
Sodium	2	99	87.9	88.6	85.8	91.8	89.7	88.9	83.3	80.9	82.0
Sodium-d			78.0	78.5	88.4	nr5	nr5	nr5	77.9	78.4	78.6
Strontium	1		0.477	0.495	0.466	0.533	0.538	0.526	0.555	0.540	0.565
Strontium-d			0.510	0.484	0.479	nr5	nr5	nr5	0.505	0.500	0.501
Sulfur	2		50.0	50.8	50.0	55.6	53.4	53.8	55.2	54.6	53.9
Sulfur-d			53.8	44.9	50.9	nr5	nr5	nr5	53.7	53.0	52.8
Thallium			<0.00005	<0.00005	<0.00005	<0.00005	<0.00005	<0.00005	<0.00005	<0.00005	<0.00005
Thallium-d			<0.00005	<0.00005	<0.00005	nr5	nr5	nr5	<0.00005	<0.00005	<0.00005
Tin			<0.001	<0.001	<0.001	<0.001	<0.001	0.003	<0.001	<0.001	<0.001
Titanium	5		0.0031	0.0040	0.0029	0.0028	0.0031	0.0026	0.0034	0.0039	0.0034
Titanium-d			0.0035	0.0027	0.0032	nr5	nr5	nr5	0.0033	0.0033	0.0032
Vanadium			0.0007	0.0008	0.0007	0.0006	0.0006	0.0006	0.0004	0.0004	0.0004
Vanadium-d			0.0003	0.0002	0.0004	nr5	nr5	nr5	0.0002	0.0003	0.0004
Zinc	5		0.045	0.049	0.044	0.044	0.045	0.041	0.038	0.040	.036
Zinc-d			0.040	0.052	0.043	nr5	nr5	nr5	0.036	0.037	0.034

* Analyzed by NORWEST LABS, Edmonton following Standard Methods 3120B Inductively Coupled Plasma (ICP) method.

[A] (Influent metal's concentration – permeate metal's concentration) / Influent metal's concentration × 100 %

[B] Influent metal's concentration / Overflow metal's concentration × 100 %

Appendix F: Calculation Procedure for Figure 4.30.

Calculation procedure for Figure 4.30 :

1. Get 24 hours operating data under different operating parameter set points, which could run the system at least 24 hours.
2. Calculate the average temperature within this 24 hours and using Equation 12 to correct the flux.
3. Delete the all backpulse data.
4. Get the average TMP and flux for each operating condition.
5. Finally all the data are summarized in Table

(Note: due to the big amount of data the raw data used for this calculation are not listed here.)

Table F.1. Average TMP and flux for different operating conditions.

15 min operating time		10 min operating time		25 min operating time	
Average TMP (kPa)	Average flux ($Lm^{-2}h^{-1}$)	Average TMP (kPa)	Average flux ($Lm^{-2}h^{-1}$)	Average TMP (kPa)	Average flux ($Lm^{-2}h^{-1}$)
16.72	25.96	22.99	33.42	14.67	20.74
12.25	20.66	17.54	28.49	9.98	16.46